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RECOVERY OF COAGULANT, NITROGEN REMOVAL, AND CARBON REGENERATION IN WASTE WATER RECLAMATION

David W. Callahan, et al

South Tahoe Public Utility District
South Lake Tahoe, California

June 1967

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B	Barnebey-Cheney Type PC
C	Darco Type DXL-0-3334

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RECOVERY OF COAGULANT, NITROGEN REMOVAL, AND CARBON REGENERATION IN WASTE WATER RECLAMATION

SOUTH TAHOE PUBLIC UTILITY DISTRICT
Project Director: David W. Callahan, Assistant Manager
Technical Consultant: Russell L. Culp
Cornell, Howland, Hayes & Merryfield

REPORT BY:

Clinton E. Smith
Clair A. Hill & Associates
Consulting Engineers
Redding, California
Engineers for
South Tahoe Public Utility District

and

Robert L. Chapman
Cornell, Howland, Hayes & Merryfield
Engineers & Planners
Corvallis, Oregon

STUDIES DURING PERIOD OF APRIL 1, 1965 TO JUNE 14, 1967

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CHAPTER I

SUMMARY

GENERAL

This report summarizes 26 months of work in waste water reclamation at the South Tahoe Public Utility District plant, South Lake Tahoe, California. The work was started in April 1965 under demonstration grants from the U. S. Public Health Service and was completed in June 1967 under a grant from the Federal Water Pollution Control Administration.

For these studies, the South Tahoe P. U. D. furnished a 2.5-mgd tertiary waste water treatment plant, including full-scale equipment for thermal regeneration of granular activated carbon and laboratory facilities, as well as financial support. The project was staffed principally by sanitary engineers from Clair A. Hill & Associates, Consulting Engineers, Redding, California (Engineers for South Tahoe P. U. D.), and from Cornell, Howland, Hayes, & Merryfield, Engineers & Planners, Corvallis, Oregon.

The purpose of the project was to evaluate the performance of the operating, advanced waste treatment plant, which included conventional biological treatment and anaerobic sludge digestion, as supplemented by chemical coagulation, mixed-media filtration, and activated carbon adsorption, with particular attention to the regeneration and reuse of granular activated carbon. In addition, investigations were to be made of coagulant recovery and reuse, and removal from the waste waters of the biological nutrients — phosphorus and nitrogen. Information gained during the studies was to be used in the proposed expansion of the plant to a total capacity of 7.5 mgd.

The objectives of the project have been fully realized. Carbon regeneration and reuse have been accomplished successfully and at reasonable costs. Methods for the recovery and reuse of lime as the coagulant have been developed and incorporated in the design of the expanded plant. An acid recovery process for recovery and reuse of alum has been shown to be practical and feasible for other situations where phosphate removal is not required. Lime was shown to be very efficient for phosphate removal. Finally, ammonia stripping at high pH values was demonstrated to be the best present method for nitrogen removal under the conditions at Tahoe, and plant-scale facilities have been designed utilizing this process.

ACTIVATED CARBON TREATMENT

At South Tahoe, additional removal of organics beyond that provided by secondary treatment is accomplished by chemical coagulation of organic particulate matter, followed by activated carbon adsorption of dissolved organics. Pilot carbon columns were operated to select the proper carbon for this use and to obtain information pertaining to the design and operation of plant-scale columns. Data have been collected on the operation of the 2.5-mgd capacity carbon columns.

Currently, carbon dosage rates have been less than 250 pounds per million gallons. The total cost of providing carbon treatment for the 7.5-mgd plant is estimated at $30 per million gallons, of which $18 is capital cost and $12 is operating cost.

COAGULANT RECOVERY

The original tertiary plant at Tahoe utilized alum as the coagulant for removal of phosphates. The high cost of alum required made it necessary to investigate methods for alum reclamation and reuse. The best method for alum recovery was by use of sulfuric acid. Unfortunately, the phosphates are also redissolved by the acid, and the cost of separating the alum from the phosphates is prohibitive. The alkaline recovery method is not applicable here, because the recovered sodium aluminate is not a good coagulant for the Tahoe waste water. However, lime is a good coagulant here, reducing phosphates to less than 1 mg/l. Spent lime was successfully recalcined to full efficiency and reused, and a method for eliminating the recycling of phosphates was developed. The cost of recalcined lime is only slightly less than that of new lime; however, the reuse of lime reduces by a factor of 2 the amount of sludge to be disposed of and therefore effects a substantial cost savings.

NITROGEN REMOVAL

Methods investigated for the removal of a second important nutrient, nitrogen, included: microbial denitrification, ion exchange, ammonia stripping, and chloramine adsorption. In combination with the lime treatment for phosphate removal, the ammonia stripping method offered several advantages over the others studied. The main requirements found for ammonia stripping are a pH of near 10.8 and sufficient air-water contact during droplet formation. The operation of a pilot ammonia stripping tower indicated that a 24-foot tower packed with a 1.5-inch by 2-inch slat packing and operated at 2.0 to 3.0 gpm per square foot and 200 to 400 cubic foot air/gallon would remove 90 to 95 percent of the ammonia-nitrogen. Since the previous addition of lime would produce a pH satisfactory for ammonia stripping, no additional chemical costs would be involved. Therefore, the estimated additional cost of ammonia stripping when used in conjunction with lime coagulation is only $15 per million gallons.

INTEGRATED PLANT DESIGN

The tertiary process developed for the South Tahoe P. U. D. will incorporate lime coagulation of secondary effluent, which, at the same time, will precipitate phosphates. Lime recalination and reuse will be employed. The effluent from the lime clarification process will pass through a stripping tower for removal of ammonia-nitrogen. This will be followed by recarbonation for pH adjustment and calcium recovery. Mixed-media filters will be used to capture $CaCO_3$ floc and provide final clarification. The final treatment phase consists of organic removal on granular activated carbon.

This process will provide an effluent with extremely low concentration of phosphorus and nitrogen, turbidity, and organic matter. All plant sludges will be incinerated to sterile, insoluble ash. The total cost of this treatment, including both operational costs and amortization of capital costs, is estimated at $150 per million gallons.

CHAPTER II

INTRODUCTION

Today, there is general recognition of the inadequacies of conventional secondary sewage treatment processes and the need for new methods which will produce effluents of superior quality.

In both the areas of waste water treatment and water purification, there is a great gap between current laboratory research findings and the practical applications and use of this knowledge in the field. The translation of the accumulated backlog of new scientific information into practical water and waste treatment processes and useful plant equipment is an engineering task which deserves more attention than it is now receiving.

Much more must be done if progress in water technology is to keep pace with that in other contemporary engineering endeavors.

The South Tahoe Public Utility District has a direct interest in advanced waste water treatment. Lake Tahoe is one of the world's most beautiful alpine lakes, and the Tahoe Basin is fast becoming one of the Nation's most popular year-around recreation areas. There is great local, State, and Federal interest in preserving the natural beauty of the area for future generations to enjoy. The remarkable clarity of the lake water is one of its principal attractions, so that pollution control is of major importance. The entire economy of the Tahoe Basin depends directly upon recreation.

No direct discharge of waste water to the lake or any watercourse within the basin has ever been permitted, and none is contemplated. Presently, all waste waters from individual and community systems are disposed of within the basin by spray irrigation or other land disposal methods. Very soon, all treated waste waters will be exported from the Lake Tahoe Basin to delay eutrophication of this body of water. In some export areas, the quality requirements for waste water are exceptionally high and, in all export areas, exceed those attainable by secondary treatment only. Future consideration of in-basin land disposal will undoubtedly require plant-scale demonstration over a period of several years of the ability to meet fully and continuously whatever quality standards may be established. It is certain that any such standards will be even higher than for export and will require nutrient removal.

The policy of the South Tahoe P. U. D. is not only to meet all present legal requirements for waste treatment but to prepare for the future by providing the maximum degree of treatment consistent with their financial means. In seeking to solve its problems, the District has become a local and National leader in pioneering new processes and methods for waste water reclamation.

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The history of this water reclamation process begins in 1961 when laboratory and pilot plant tests of chemical coagulation, mixed-media (or coarse-to-fine) filtration with polyelectrolyte filter aid, and granular carbon adsorption were made at Salem, Corvallis, and Philomath, Oregon. Following this, pilot plant studies were made on a larger scale at South Tahoe. This treatment not only increased removals of BOD, COD, suspended solids, bacteria, and other substances which are partially removed by conventional sewage treatment, but also removed color, odor, colloids, viruses, phosphates, ABS, and other organic chemicals which are relatively unaffected by secondary plants. The aesthetic and technical improvements achieved in effluent quality are quite remarkable.

In 1963, a 2.5-mgd tertiary plant was designed for the South Tahoe P.U.D. incorporating these processes, plus facilities for thermal regeneration of granular activated carbon for flows up to 10 mgd. This plant was built in 1964 and 1965 and was placed into operation during the summer of 1965.

In April 1965, further laboratory, pilot plant, and full-scale plant studies were initiated at Tahoe with demonstration grant funds. These studies include recovery and reuse of coagulant, nitrogen removal, and data collection in connection with full-scale carbon regeneration and reuse.

In 1966, plans and specifications were prepared to expand the capacity of the entire South Tahoe plant from 2.5 to 7.5 mgd. Facilities were planned for the recovery and reuse of lime as the coagulant and for the incineration of all sludge produced. These improvements are now under construction. At the present time, an ammonia stripping tower for nitrogen removal is being designed with a capacity of 7.5 mgd. It is expected that all facilities for the expanded plant will be placed into operation early in 1968, together with the pump stations and pipelines necessary to export the reclaimed water to Alpine County, California, where it will be used to fill a recreational lake and for irrigation of grassland.

This report summarizes the results of the 26 months of demonstration grant studies at the South Tahoe P. U. D., and tells how this information was used in process selection and refinement, and in the design of the 7.5-mgd plant, which is now under construction.

CHAPTER III

THE ORIGINAL SOUTH TAHOE PUBLIC UTILITY DISTRICT WATER RECLAMATION PLANT

GENERAL

A tertiary sewage treatment process has been developed(2)(3)(4) and a 2.5-mgd plant constructed at the South Tahoe (California) Public Utility District. This water reclamation plant, which was placed into operation in June 1965, treats the effluent from a conventional activated sludge plant. The tertiary process, Figure 1, consists of coagulation of secondary effluent with liquid alum, followed by series filtration through two mixed-media separation beds. The effluent from the separation beds flows upward through two carbon columns operated in parallel.

SEPARATION BEDS

The separation beds are 10 feet in diameter, 38 feet long, and are operated at a surface loading rate of 5.0 gpm/square foot. These tanks lie horizontally and contain a 3-foot depth of graded media supported on a bed of graded gravel. This media, which was developed by the Neptune-Microfloc Corporation, consists of several different materials, each with a different size and specific gravity. By proper selection of these materials, a filter bed can be obtained which is nearly uniformly graded from coarse to fine, from 10-mesh to 80-mesh, in the direction of filtration. With this distribution of particle sizes, penetration of suspended materials is possible, permitting complete utilization of the entire filter bed depth, rather than only the top few inches of media, as is the case with conventional sand filters. The gradation of the filter media is maintained during and after the hydraulic backwash of the beds. The advantages of this type of filter media over a typical, rapid sand filter are many:

1. The surface-loading rate on the filters can be doubled or tripled, and the length of filter runs between backwash periods can be considerably extended.

2. With the system of gradation employed, the finest filter media can be reduced to 0.15 mm as opposed to the 0.45 mm in a sand bed. This allows the production of a higher quality effluent.

3. Polyelectrolytes can be employed to control the penetration of suspended material into the bed. Thus, the storage capacity of the filter bed can be greatly increased without floc breakthrough.

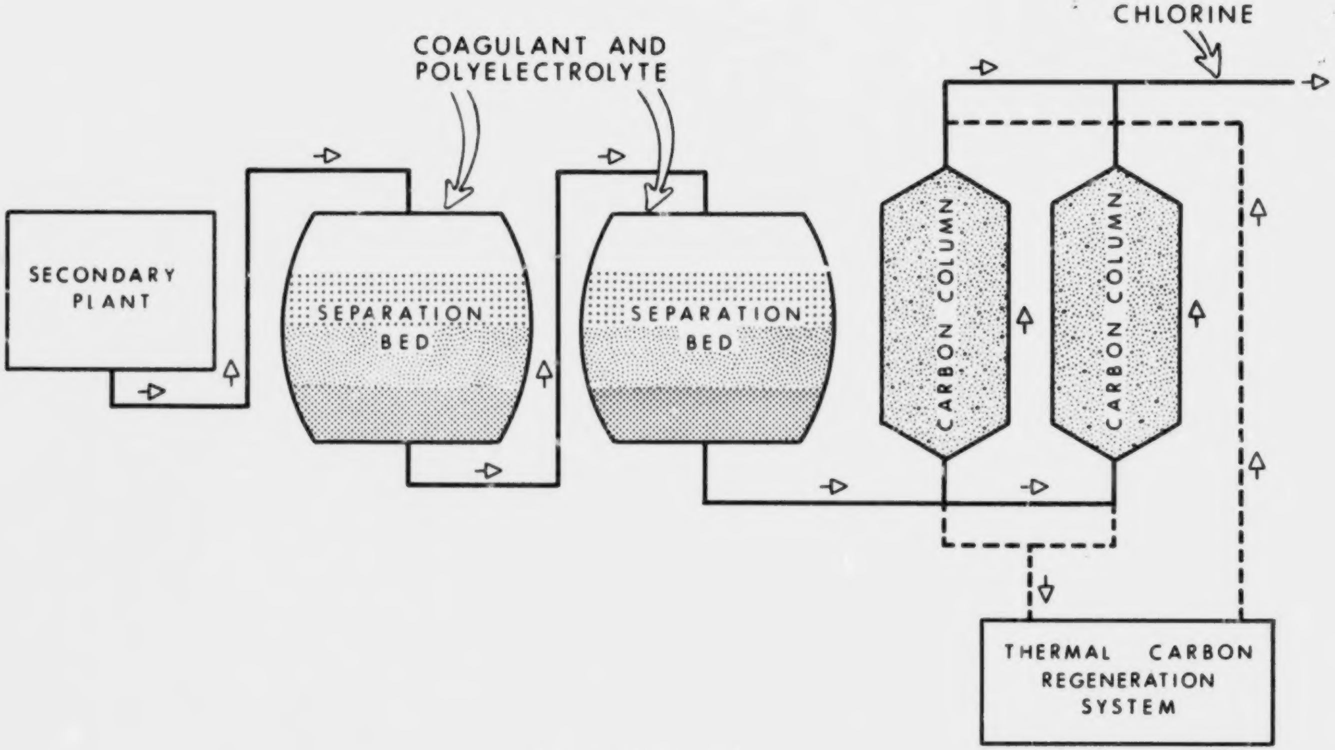
COAGULANT AND
POLYELECTROLYTE
CHLORINE
SECONDARY PLANT
SEPARATION BED
SEPARATION BED
CARBON COLUMN
CARBON COLUMN
THERMAL CARBON REGENERATION SYSTEM

FIGURE 1
FLOW THROUGH ORIGINAL TERTIARY PLANT

Several unique items were incorporated into the design of the South Tahoe tertiary treatment plant.

1. The filtration process uses two separation beds in series to utilize the mixed-media beds to their fullest extent.

2. Series backwash of the two beds reduces the backwash water requirements, lessens the capital investment, and shortens the total downtime for backwash.

This clarification process serves not only to remove particulate matter from the final effluent, but to reduce the phosphate concentration as well. This is accomplished by adsorption of the phosphates on the alum floc. In addition, the process serves as a pretreatment for the carbon columns. Problems of physical fouling of the carbon and excessive head loss development across the carbon columns are thereby minimized.

CARBON COLUMNS

The second basic treatment step involves the removal of dissolved organic material by adsorption. The effluent from the separation beds flows upward through two columns of activated carbon operated in parallel. Each column, as illustrated by Figures 2 and 3, is 12 feet in diameter and has an effective carbon depth of 14 feet.

The water flows into the column through a series of eight, 12-inch-diameter, stainless steel well screens located near the bottom. In a similar fashion, the water leaves the column through well screens near the top of the column. Spent carbon is periodically withdrawn from the bottom of the column, while replacement of regenerated carbon is added at the top. Thus, a countercurrent operation is established, which makes possible complete exhaustion of the carbon before withdrawal, as well as the final passage of the effluent over freshly regenerated carbon. The water flow is occasionally reversed to downflush any particulate matter trapped in the lower sections of the carbon bed.

Head loss through each carbon column is continuously monitored and recorded. Flow rate is manually controlled by a valve on the discharge line and indicated on a flow meter.

CARBON REGENERATION SYSTEM

In order to make the process economically feasible, regeneration of the activated carbon is necessary. Spent carbon is transferred hydraulically from the carbon columns to storage and dewatering bins. A screw conveyor feeds the carbon at a controlled rate to the top hearth of the regeneration furnace. within the furnace, the carbon is heated to near 1,700°F, and steam is added to obtain more uniform distribution of heat in the furnace. Organic impurities

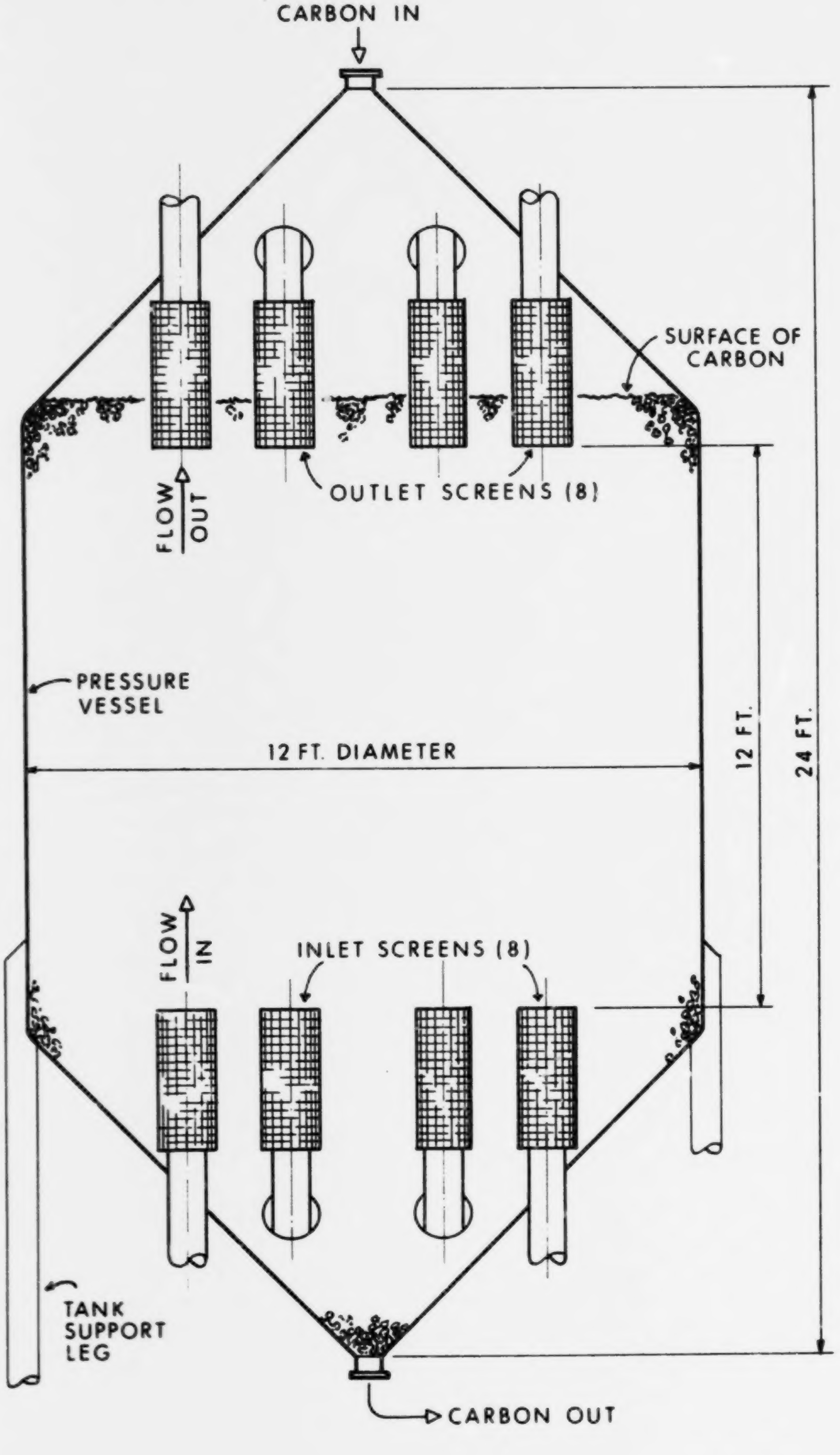
CARBON IN
SURFACE OF CARBON
FLOW OUT
OUTLET SCREENS (8)
PRESSURE VESSEL
12 FT. DIAMETER
12 FT.
24 FT.
FLOW IN
INLET SCREENS (8)
TANK SUPPORT LEG
CARBON OUT

FIGURE 2
SECTION THROUGH CARBON COLUMN

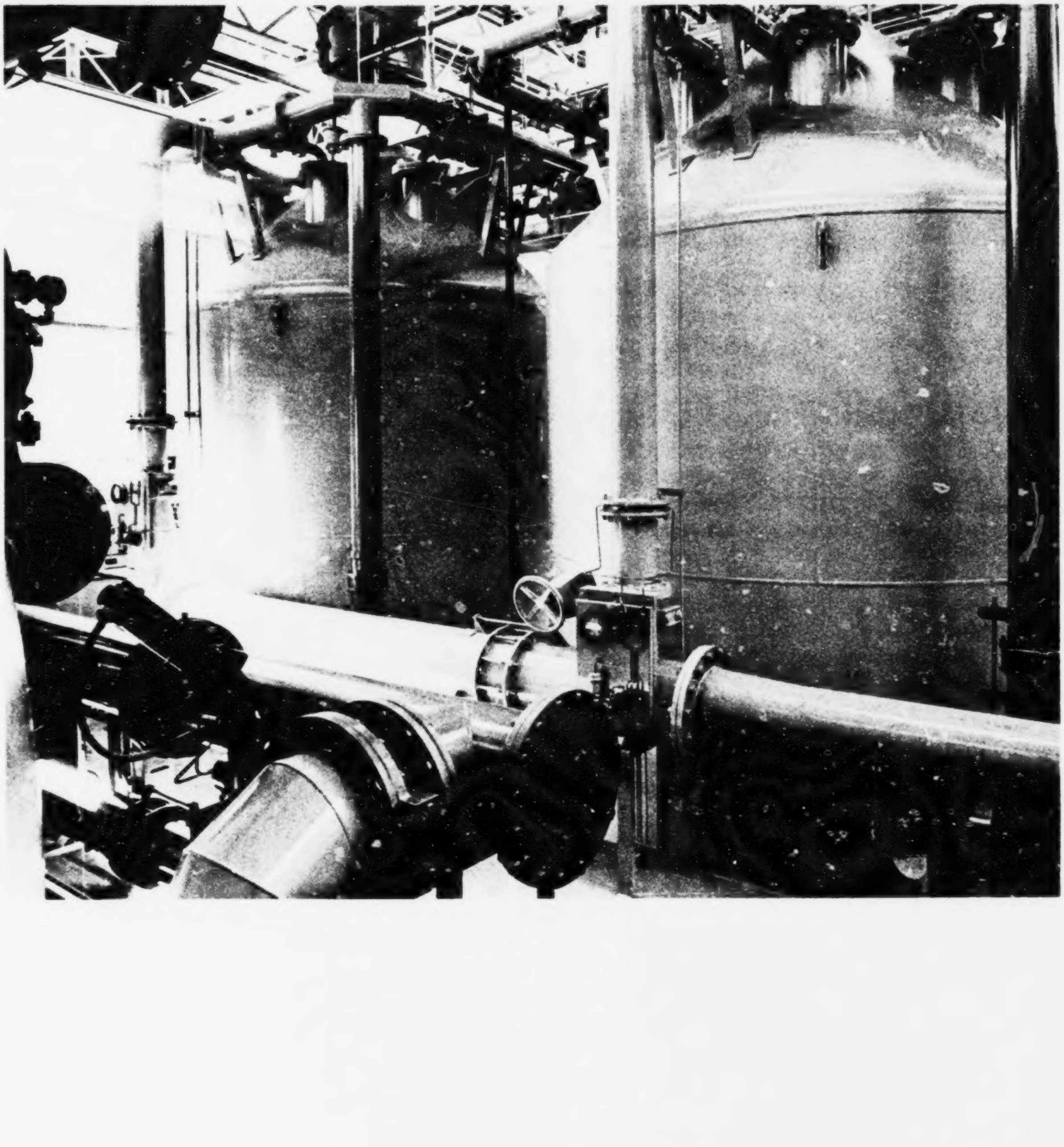

FIGURE 3
CARBON COLUMNS

in the carbon are thus volatilized and driven off in gaseous form. Exhaust emissions are controlled by a water scrubber. An afterburner is also available for use as necessary, but has not been needed to date. The carbon is discharged from the furnace to a quench tank and then pumped to a storage and defining tank prior to returning to the top of the carbon columns. A picture of the carbon handling and regeneration system appears in Figure 4.

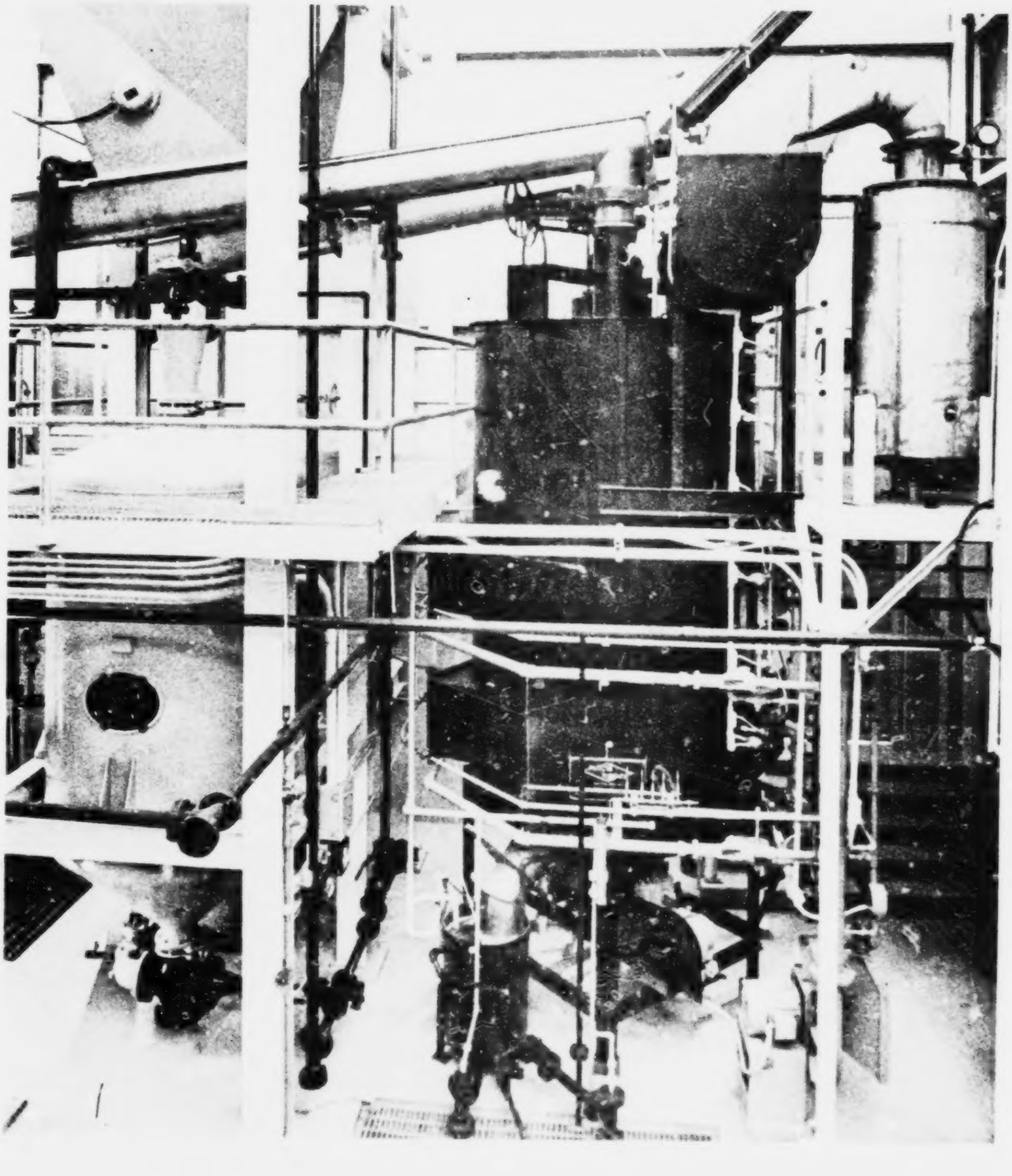

FIGURE 4
CARBON REGENERATION EQUIPMENT

TREATMENT EFFECTIVENESS

The effectiveness of the treatment processes can best be conveyed by means of Table 1. Therein is listed average values for the quality of water after various stages of treatment as measured by important parameters. A visual presentation of water quality appears in Figure 5.

Additional detailed information on the operational results of the activated carbon treatment is presented in a later section of this report.

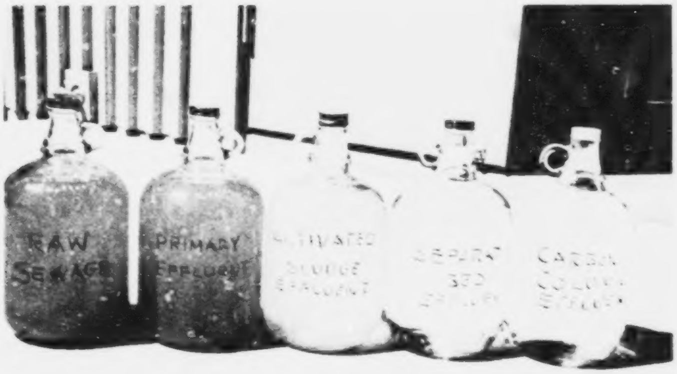

FIGURE 5
PROCESS EFFLUENT STREAMS-EXISTING
SO. TAHOE P.U.D. WATER RECLAMATION PLANT

Table 1

WATER QUALITY AT VARIOUS POINTS IN ORIGINAL PROCESS

	Raw Sewage	Secondary Effluent	Separation Bed Effluent	Carbon Column Effluent
BOD, mg/l	200–400	20–100	<1	<1
COD, mg/l	400–600	80–160	20–60	3–25
Total Organic Carbon, mg/l	–	–	7–18	1–8
ABS, mg/l	2.0–4.0	1.1–2.9	0.5–2.9	<.01–0.50
PO_4, mg/l as PO_4	–	25–30	0.1–1	0.1–1
Color, units	–	–	20–30	<10
Turbidity, units	–	30–70	<0.1–1.0	<0.5
Nitrogen				
Organic N, mg/l as N	10–15	4–6	2–4	1–2
Ammonia N, mg/l as N	25–35	25–35	25–35	25–35
NO_3 and NO_2, mg/l as N	0	0	0	0
Unchlorinated				
Coliforms, MPN/100 ml	–	2,400,000	9,300	11,000
Fecal Coliforms, MPN/100 ml	–	150,000	930	930
Chlorinated				
Coliforms, MPN/100 ml	–	–	8.6	<2.1
Fecal Coliforms, MPN/100 ml	–	–	<2.1	<2.1

(Samples for analysis for virus were obtained from separation bed effluent and carbon column effluent, and were sent to the Robert A. Taft Sanitary Engineering Center. All samples were reported negative.)

CHAPTER IV

LIST OF PROJECT PERSONNEL

Following completion of the 2.5-mgd South Tahoe Public Utility District Water Reclamation Plant, demonstration grant research and data collection were instigated. The following personnel have performed or directed the work described in the remainder of this report:

	Period With Demonstration Grant Project
W. F. Priday, Former Project Director Former Manager — South Tahoe P.U.D.	Mar 1965 to Jul 1966
David W. Callahan, Project Director Assistant Manager, South Tahoe P.U.D.	Jul 1965 to Jun 1967
Harlan E. Moyer, District Engineer and Acting General Manager Clair A. Hill & Associates, Consulting Engineers for South Tahoe P.U.D.	Apr 1965 to Jun 1967
Russell L. Culp, Technical Consultant Principal — Cornell, Howland, Hayes & Merryfield	Mar 1965 to Jun 1967
Alfred Slechta, Sanitary Engineer Clair A. Hill & Associates, Consulting Engineers for South Tahoe P.U.D.	Apr 1965 to Jun 1967
Gordon L. Culp, Sanitary Engineer Cornell, Howland, Hayes & Merryfield	Apr 1965 to May 1966
William Chew, Laboratory Technician Graduate Student, University of California	Jun 1965 to Sep 1965
John F. Steward, Civil Engineer Cornell, Howland, Hayes & Merryfield	Sep 1965 to Sep 1966
Clinton E. Smith, Sanitary Engineer Clair A. Hill & Associates, Consulting Engineers for South Tahoe P.U.D.	Aug 1966 to Jun 1967
Robert L. Chapman, Sanitary Engineer Cornell, Howland, Hayes & Merryfield	Sep 1966 to Jun 1967

CHAPTER V

ACTIVATED CARBON TREATMENT

GENERAL

Activated carbon polishing of coagulated and filtered secondary effluent is one of the main features of the South Tahoe P. U. D. Water Reclamation Plant. This section contains a review of activated carbon treatment, the results obtained from the plant-scale carbon treatment system at South Tahoe, and the results of associated pilot carbon column research.

THEORY

Activated carbon removes organic impurities from water by the process of adsorption. Adsorption may be defined(5) as the phenomenon whereby molecules adhere to a surface with which they come in contact. The fact that activated carbon has extremely large surface area per unit of volume makes it an efficient adsorptive material. The activation of carbon produces many pores in the carbon particles, and it is the area of the walls within these pores that accounts for nearly all of the surface area of the carbon particles.

Upon contact with a water containing soluble organic materials, the activated carbon selectively removes these organic materials by adsorption. Activated carbon can be made from over 30 types of raw materials.(6) The carbons used in water treatment are usually derived from wood or coal. Carbons made from bituminous coal are hard and dense and can be pumped in a water slurry without any appreciable deterioration.(7) Carbons made from wood are usually not hard enough to be handled hydraulically. The physical strength of the carbon must be great enough to withstand the repeated handling procedures associated with the regeneration processes.

Thermal regeneration is the only known practical means of restoring virgin activity to carbon which has been used in water treatment.(8) The temperature of the spent carbon is raised to a point where the adsorbed organic materials are volatilized and a minimum amount of burning of the carbon occurs. Regeneration of granular carbon is generally accomplished in multiple-hearth furnaces or rotary kilns.

LITERATURE REVIEW

The use of carbon for purposes other than fuel is reported as early as 1550 B.C., when it was used for medicinal purposes.(6) The discovery of the phenomenon of adsorption, as now understood, is generally attributed to Scheele,

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who, in 1773, described experiments on gases exposed to carbon.(6) The earliest documented use of carbon for removal of impurities from solutions is 1785, when Lowitz observed that charcoal would decolorize many liquids.(6) The first industrial application came a few years later when wood char was used in a refinery to clarify cane sugar.(6)

During the last 40 years, activated carbon has been used extensively in water treatment for the removal of taste- and odor-producing substances. During this period, powdered carbon was used far more than granular activated carbon beds for water treatment. Since regeneration of granular activated carbon has been proven to be economical, granular activated carbon beds have been used in several instances(7,8,9,10) for the control of tastes and odors in domestic water supplies. The use of granular carbon beds has two important advantages over the use of powdered carbon.(9) More efficient use can be made of carbon in the bed form than in the slurry form, and there is no need to continually check for the proper dose with carbon beds. Hassler(6) reports that for 137 municipalities using activated carbon for taste and odor control, the average carbon dose was 23 pounds per million gallons, with a range of 4 to 429 pounds per million gallons.

The U.S. Public Health Service feels that the adsorptive process is the most promising of currently known methods for the further treatment of secondary sewage effluents.(11) Of course, the treatment of these effluents requires larger carbon doses than those normally used in water treatment. Because of the large quantities of carbon required, efficient regeneration of the carbon is a requirement for an economical waste water treatment system. To date, the use of granular activated carbon beds has received the most study for adsorption of organic materials from sewage effluents. Difficulty with the recovery and regeneration of powdered carbon has, so far, limited its application to sewage effluents.

Morris and Weber(12) have studied the adsorption process applied to materials commonly found in sewage effluents. They found that the rate of adsorption is a linear function of the square root of concentration of the material being adsorbed. Thus, the process favors the removal of trace quantities. They found that the rate of adsorption varied as the inverse square of the diameter of the carbon particles. As pointed out by Flentje and Hager,(7) crushing the carbon particles to produce smaller particles enhances the rate of adsorption by exposing more entrances to the carbon pores. Because the carbon particle size primarily affects the rate of adsorption and not the capacity of the carbon, the difference in performance of columns containing different size carbons decreases as the contact time increases.(13)

Morris and Weber(12) also found that the configuration of the adsorbate molecule affects the rate of adsorption. Molecules with highly branched structures are removed much more slowly than those of identical molecular weight but with a configuration that permits coiling or compactness. They found the rate of adsorption of ABS to increase with decreasing pH

Joyce et al(14,15) studied the treatment of a sand-filtered secondary effluent with granular activated carbon beds on a pilot scale. They found the required carbon dose to be a function of contact time for hydraulic loading rates in the range of 2.5 to 10 gpm/square foot. The carbon dose varied from 500 to 830 pounds per million gallons of filtered secondary effluent treated. The chemical oxygen demand (COD) of the filtered secondary effluent was reduced from 50 to 70 mg/l to 12 to 20 mg/l by the carbon treatment. The remaining COD could not be reduced by increasing the carbon column contact time from 15 minutes to 37 minutes. The ABS content of the carbon column effluent was used as a control. When ABS reached 0.5 mg/l in the carbon column effluent, spent carbon was regenerated. The spent carbon removed for regeneration had adsorbed about 4 percent ABS by weight. The spent carbon was regenerated at 1,740°F, using a 34-percent steam, 66-percent air atmosphere. Repeated regeneration (15 cycles) of the carbon reduced the total surface area, as reflected by the decrease in iodine number from 942 for the virgin carbon to 782 at the end of the test. Some of the finer pores in the carbon were widened, and the carbon showed a marked increase in ash, from 5.4 percent to 13.2 percent. The greatest portion of the ash was found to be acid soluble; thus, it was concluded that, if it should be necessary, the ash buildup in the regenerated carbon could be prevented by treatment of the carbon with acid following regeneration. Carbon losses averaged 4.6 percent per cycle of regeneration.

After 15 regeneration cycles, a column of regenerated carbon was operated in parallel with a column of virgin carbon. The column of regenerated carbon passed 32 percent less flow prior to ABS breakthrough than did the column of virgin carbon. Based upon their pilot plant data, Joyce et al(14) estimated the total cost of activated carbon treatment of secondary effluent to be $30 per million gallons for a 100-mgd plant, $70 per million gallons for a 10-mgd plant, and $130 per million gallons for a 1-mgd plant.

An 0.3-mgd activated carbon pilot plant has been operated at Pomona, California by the Los Angeles County Sanitation District.(16) This pilot plant treats activated sludge plant effluent in a four-stage, downflow column of granular activated carbon. Included in the plant is a furnace for carbon regeneration.

After one year of operation, the average dosage was 250 pounds of carbon per million gallons of water treated. It was also noted that the capacity of the carbon for COD could not be completely exhausted, which led to the conclusion that the column was supporting a biological growth.

The carbon used is a special grade, 16 x 40-mesh carbon, with an initial iodine number of nearly 1,100. The iodine number has steadily dropped with each regeneration to a value of about 788. The efficiency of the carbon did not reflect the dropping iodine numbers.

The cost of carbon treatment in a 10-mgd plant of this type, based on an estimated carbon loss of 10 percent, is expected to be $80 per million gallons.

Davies and Kaplan(17) studied a pilot-scale adsorption system for sewage effluents using powdered carbon. Powdered carbon doses of 2,000 to 5,000 pounds per million gallons of secondary effluent were used. The powdered carbon was mixed with the effluent, and the carbon was then flocculated with the aid of a polyelectrolyte and separated from the effluent in a sludge-blanket settling system. About 90 percent of the equilibrium capacity of the carbon was reached in a 5-minute mixing period. The carbon-treated effluent COD was about 5 mg/l. The spent powdered carbon was regenerated in a steam atmosphere at 750°F, with a 1-hour detention in the furnace. The estimated cost of the powdered carbon treatment was $67 to $79 per million gallons for a 10-mgd plant, or about the same as the estimated cost of the granular carbon system.(14) Although powdered carbon can be very successfully regenerated in the laboratory, as shown by Davies and Kaplan,(17) full-scale regeneration systems for powdered carbon have not been so successful. Sugar refineries attempted regeneration of powdered carbon for many years but finally abandoned the process because of the irregular recoveries (60 to 90 percent) of the regeneration process.(6) On the other hand, full-scale regeneration systems for granular carbon are widely used in the sugar industry with carbon losses of 5 percent used for design purposes. Five-percent losses have been independently reported for granular regeneration systems used in municipal water treatment.(8) Although the data of Davies and Kaplan(17) indicate some potential advantages of a powdered carbon system (lower carbon regeneration temperature, effective treatment of unfiltered secondary effluent), the practicality of a full-scale powdered carbon recovery and regeneration system for waste water treatment has yet to be demonstrated.

The economics of the granular carbon adsorption system, applied to waste water, are dependent upon the efficiency of the regeneration process. The cost of the carbon added to replace that lost in the regeneration process may constitute more than 50 percent of the total cost of carbon treatment.(15) The available cost estimates for granular activated carbon treatment of waste water are based only on pilot-scale plants.

ACTIVATED CARBON COLUMN STUDIES — MAIN CARBON COLUMNS

The main columns of activated carbon, as described earlier in this report (Figures 2 and 3), were each charged with 25 tons of 8 x 30-mesh activated carbon (Pittsburgh Activated Carbon Company, Type SGL). Sampling taps are provided on the influent and effluent headers. A 24-hour composite sample of the influent to the carbon columns is collected by means of a timer and solenoid valve. In addition, grab samples of both the carbon column influent and effluent are collected daily.

Analysis and Results

All of the samples collected were anlayzed for apparent alkyl benzene sulfonate (ABS), chemical oxygen demand (COD), and total organic carbon (TOC). Analyses for ABS and COD were performed by study personnel, and the TOC analyses were performed by the Water Pollution Control Administration, Robert A. Taft Sanitary Engineering Center, Cincinnati, Ohio.

Figures 6 through 8 show the curves obtained for the removal of ABS, COD, and TOC in the main carbon columns and the associated carbon dose at each carbon withdrawal. To date, approximately 366 million gallons of separation bed effluent have been passed through the carbon columns at an average carbon dose of 405 pounds of carbon per million gallons, as indicated in Table 2. The average residence time in the columns was 18 minutes. The initial high carbon dosages were due to the removal of carbon, which was only partially spent but was contributing to excessive head loss.

Table 2

CARBON DOSAGE

Date		Volume Passed (mg)	Total Volume (mg)	Carbon Removed (lb x 10^3)	Total Carbon Removed (lb x 10^3)	Carbon Dose (lb/mg)	Average Carbon Dose (lb/mg)
From	To						
Start	9/16/65	55.3	55.3	49.4	49.4	894	894
9/16/65	10/7/65	17.5	72.8	10.4	59.8	594	821
10/7/65	10/15/65	8.4	81.2	13.4	73.2	1,590	902
10/15/65	2/1/66	31.3	112.5	7.9	81.1	252	721
2/1/66	8/1/66	20.4	132.9	6.0	87.1	294	656
8/1/66	9/12/66	52.5	185.4	16.9	104.0	322	561
9/12/66	10/21/66	28.5	213.9	6.0	110.0	211	517
10/21/66	11/7/66	14.0	227.9	10.1	120.1	721	527
11/7/66	1/4/67	38.3	266.2	9.1	129.2	238	486
1/4/67	3/17/67	49.6	315.8	10.1	139.3	204	441
3/17/67	4/21/67	50.7	366.5	9.2	148.5	182	405

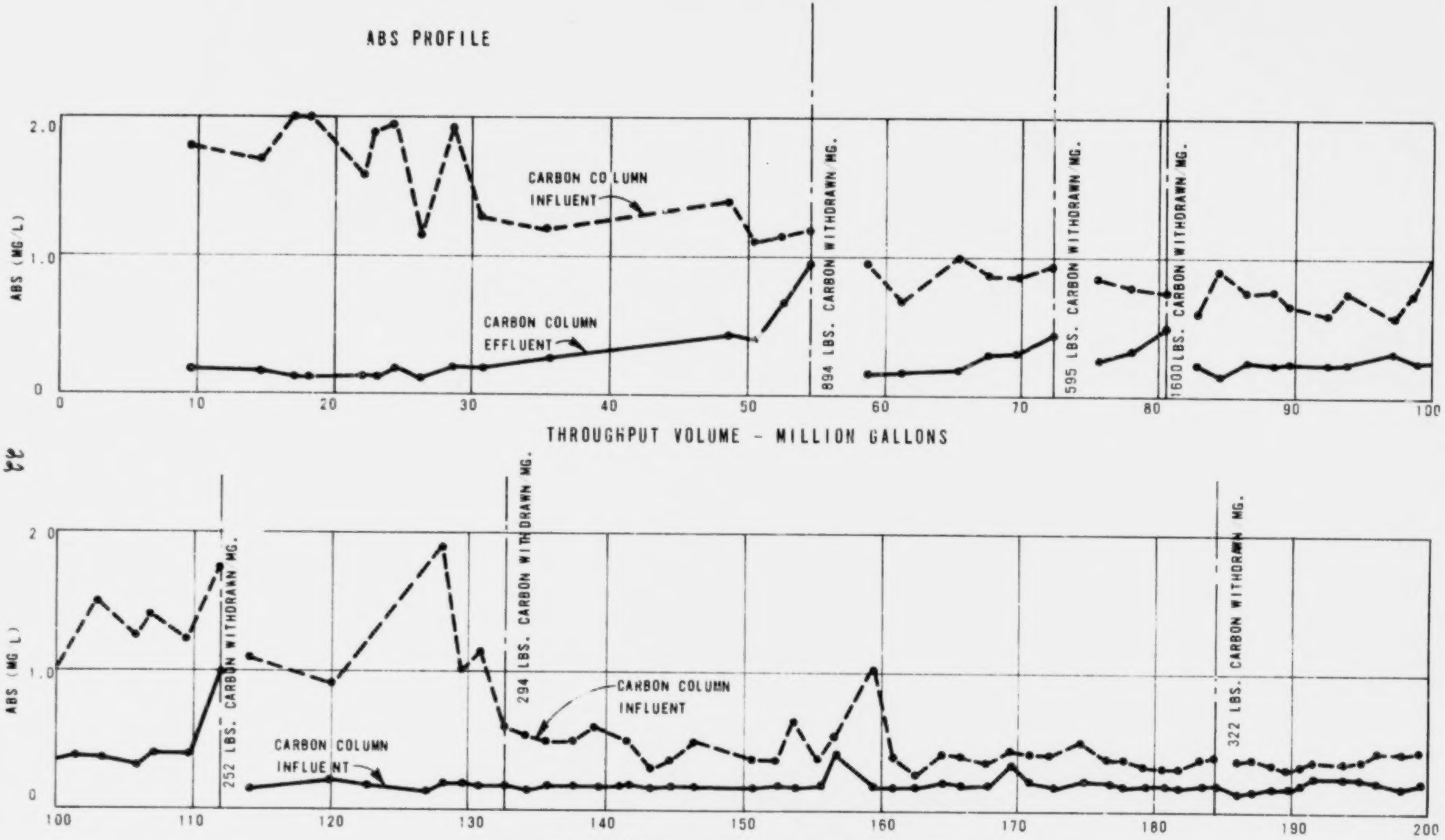

FIGURE 6

MAIN CARBON COLUMN OPERATING DATA

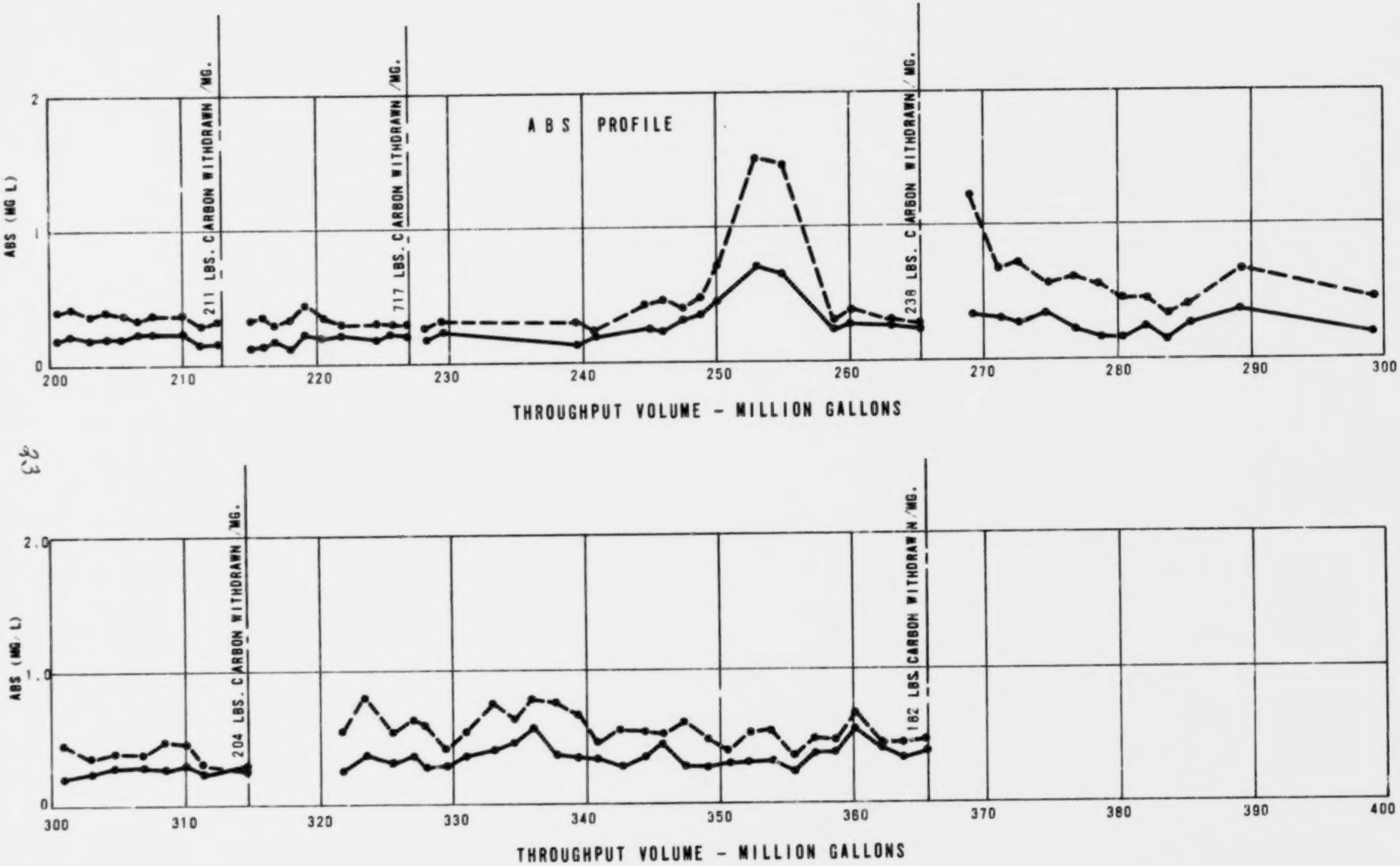

FIGURE 6

MAIN CARBON COLUMN OPERATING DATA

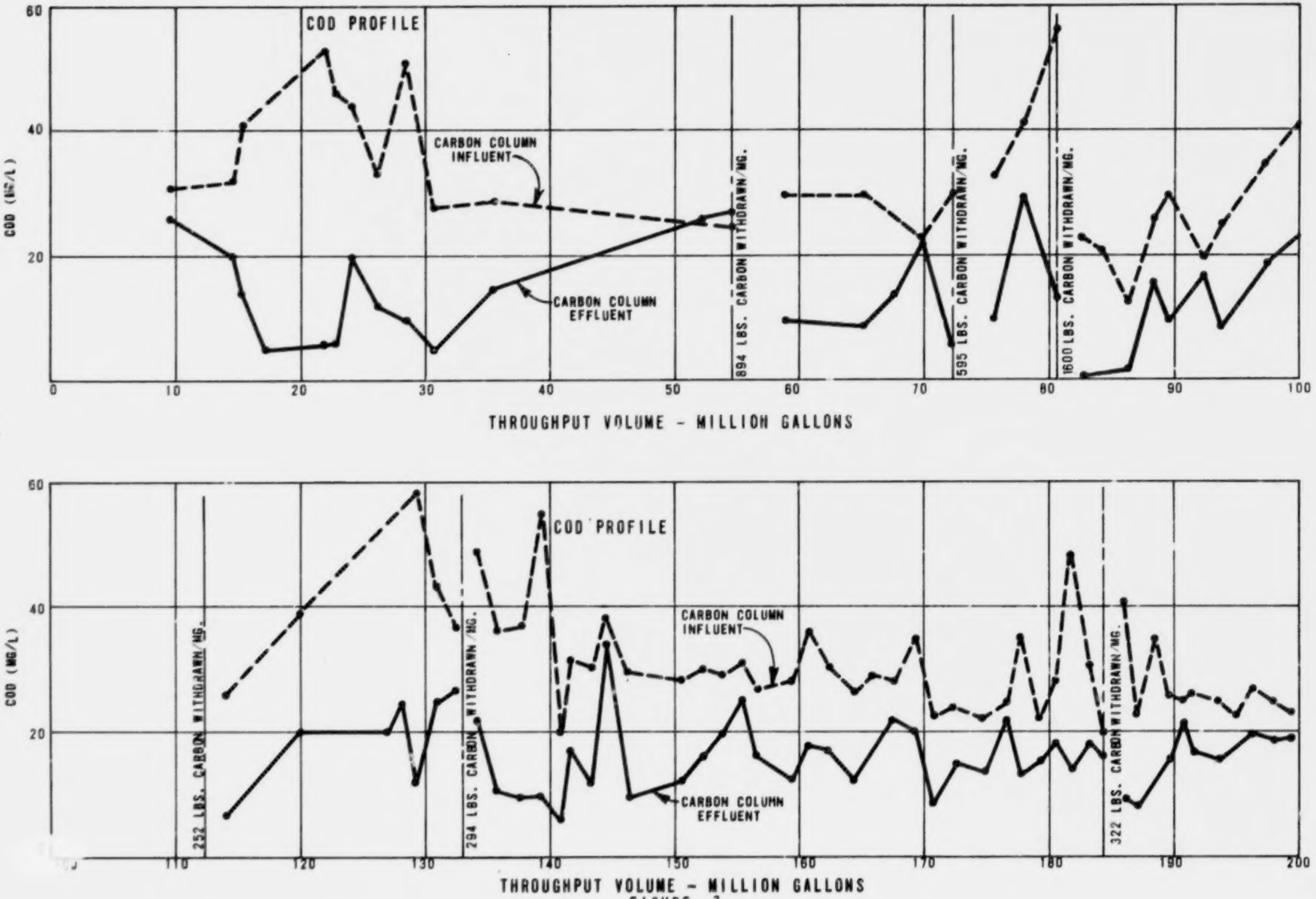

FIGURE 7

MAIN CARBON COLUMN OPERATING DATA

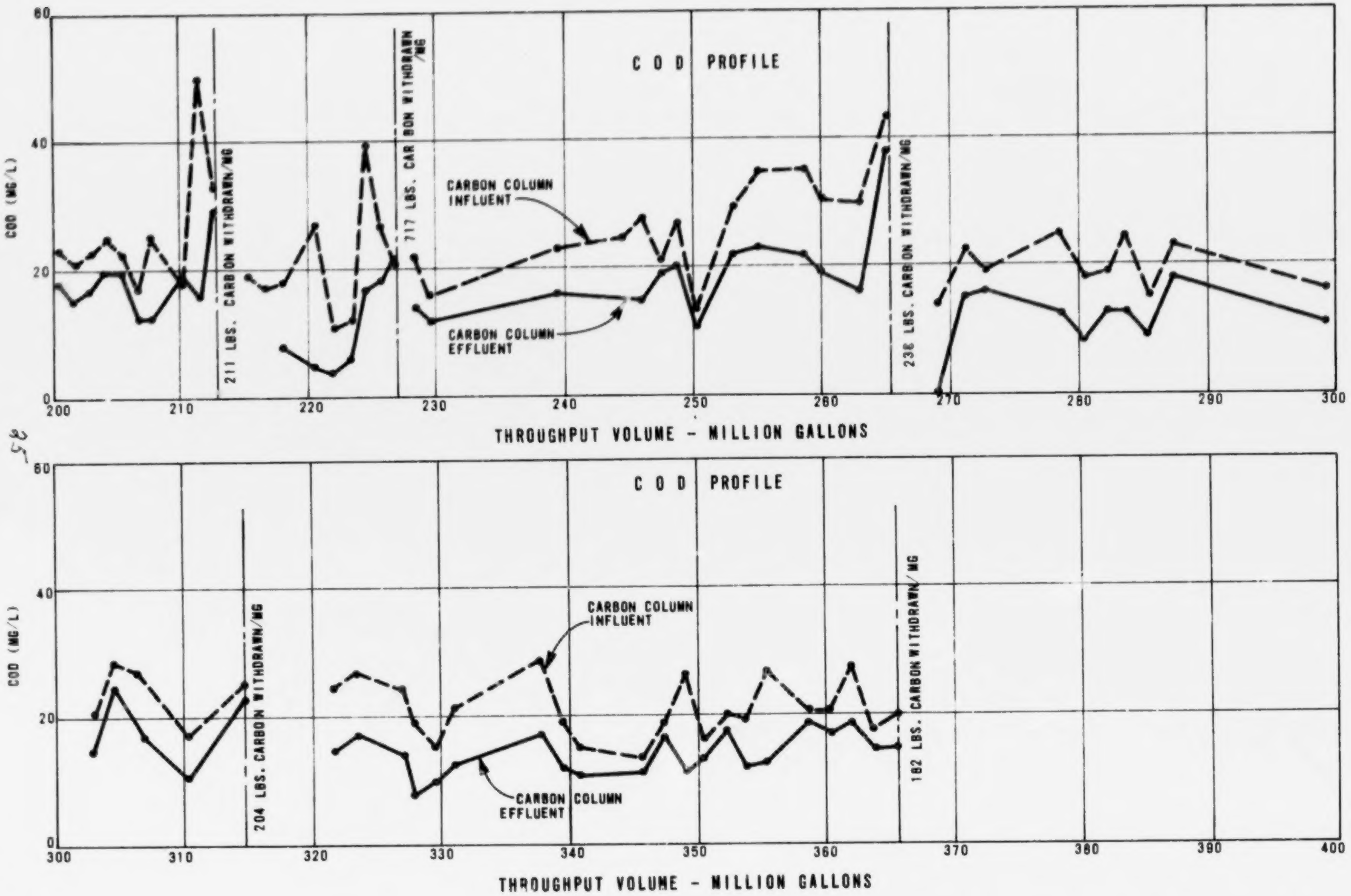

FIGURE 7

MAIN CARBON COLUMN OPERATING DATA

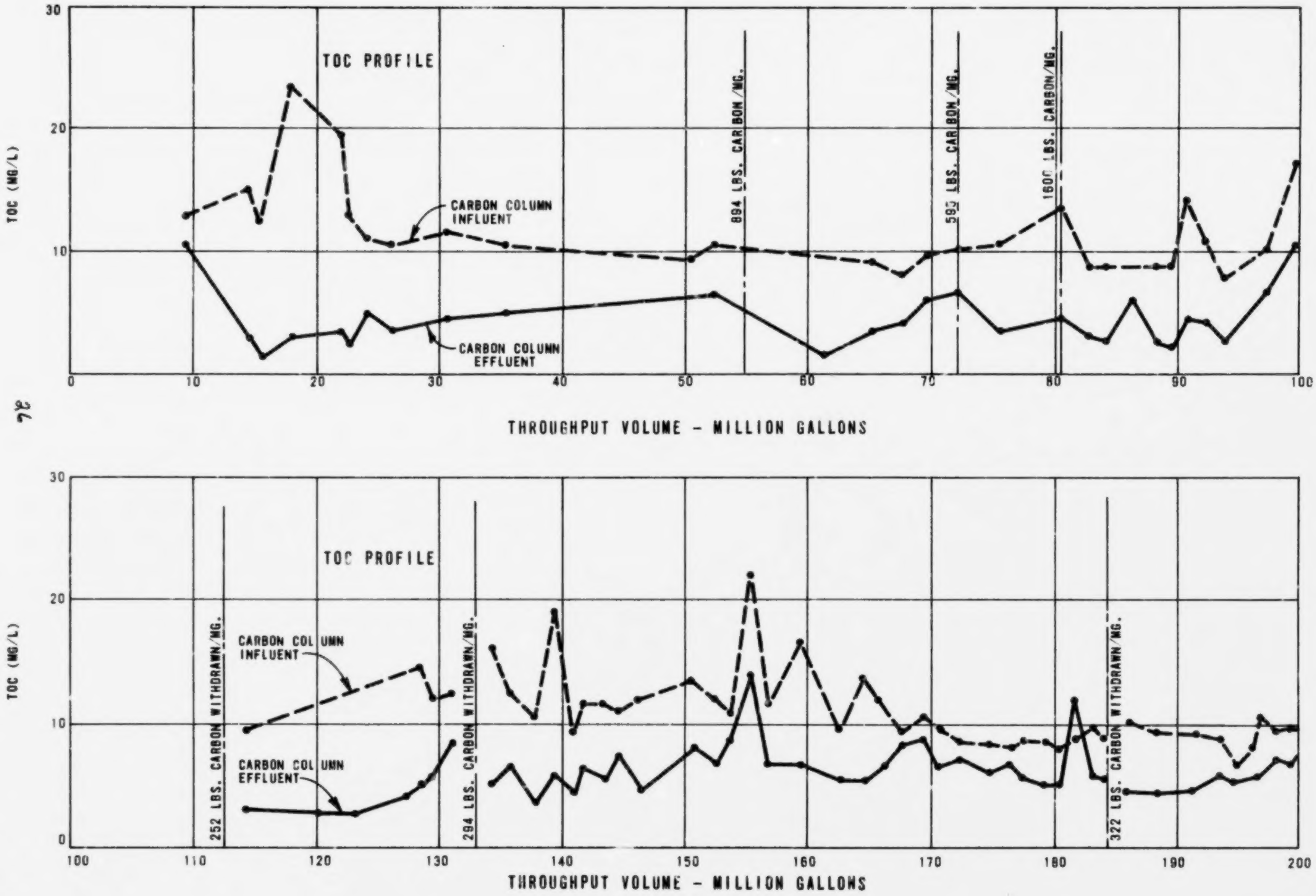

FIGURE 8

MAIN CARBON COLUMN OPERATING DATA

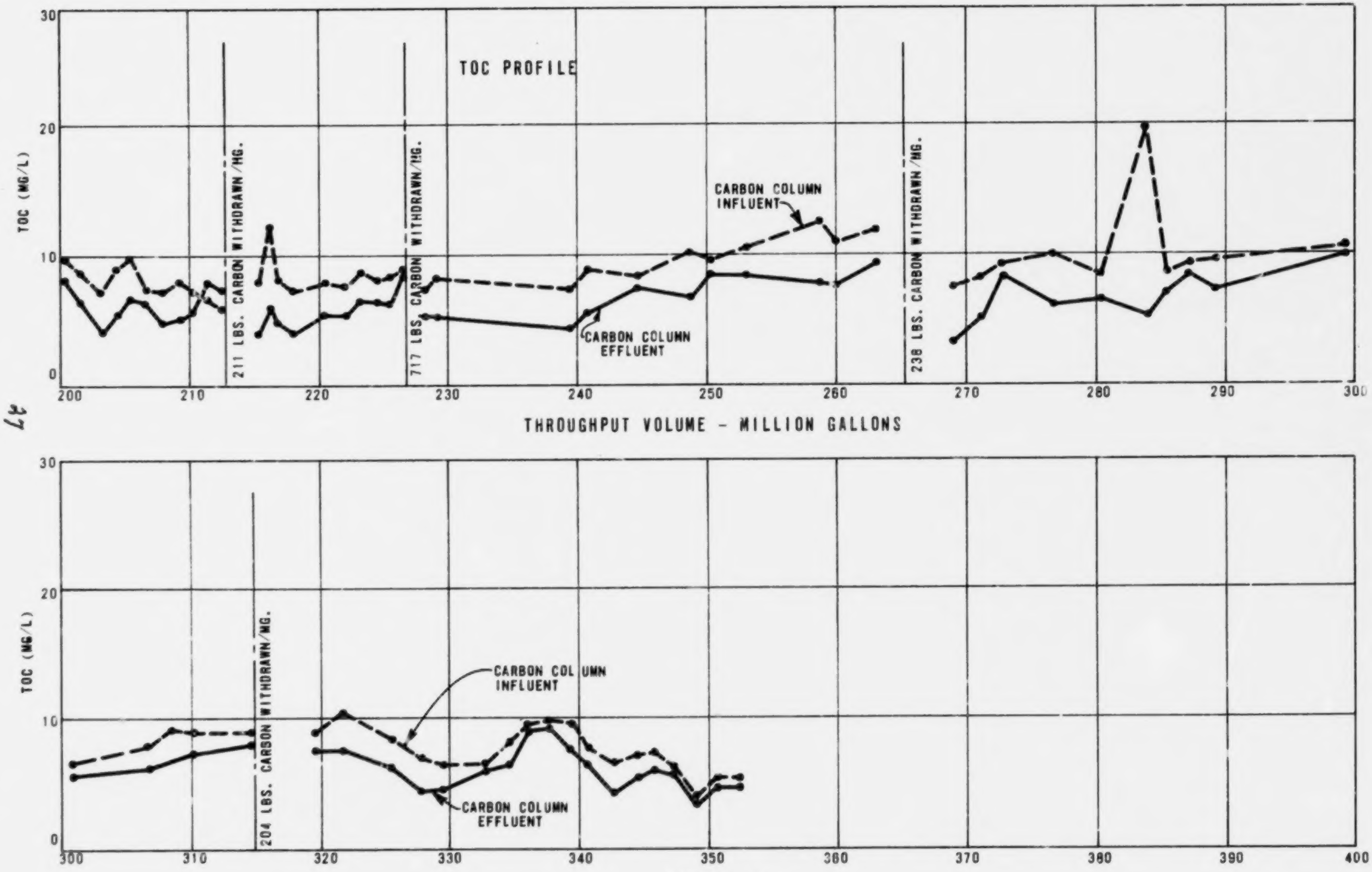

FIGURE 8

MAIN CARBON COLUMN OPERATING DATA

Operational Problems

Periodically, the head loss in one or both of the carbon columns has increased to such an extent that the plant flow is reduced. These periods of high head loss usually coincide with or follow an increase of turbidity in the column influent. During these times, it is necessary to reverse the flow through the columns and flush out the trapped particulate matter at the bottom, which has been filtered by the carbon bed. It has been found that daily down-flushing of the carbon column will effectively keep the carbon free of particulate matter and eliminate problems of high head loss.

CARBON REGENERATION

Description of Regeneration System

A flow sheet of the plant-scale carbon regeneration system is shown on Figure 9. When the carbon column effluent reaches the minimum effluent standards, spent carbon is removed from the column. These minimum water quality standards are as follows:

ABS	0.5 mg/l
COD	20 mg/l
Color	10 units
TOC	7.5 mg/l

Spent carbon is withdrawn and transferred in a slurry to the spent carbon drain and feed tanks by means of unchlorinated plant effluent. The carbon feed tanks are equipped with a screened drain to facilitate dewatering the carbon. After the carbon is dewatered, the carbon is fed to the furnace by a stainless steel screw conveyor.

Carbon regeneration takes place in a six-hearth, propane-fired Pacific furnace, equipped with an afterburner and scrubber. The carbon flows downward through the six hearths in series. The hearths are numbered 1 to 6 from top to bottom. The capacity of the furnace is 6,000 pounds dry carbon per day. Burners are located on hearths 4 and 6 and are equipped with automatic temperature controllers. In addition, steam can be supplied to hearths 4 and 6.

The regenerated carbon discharges into the quench tank. The carbon is pumped from the quench tank by diaphragm pumps to the carbon defining tanks. The carbon is washed and is then moved hydraulically to the top of the carbon columns.

Under normal operation, a surplus carbon inventory is maintained. This permits bypassing the carbon columns for only a very short period of time,

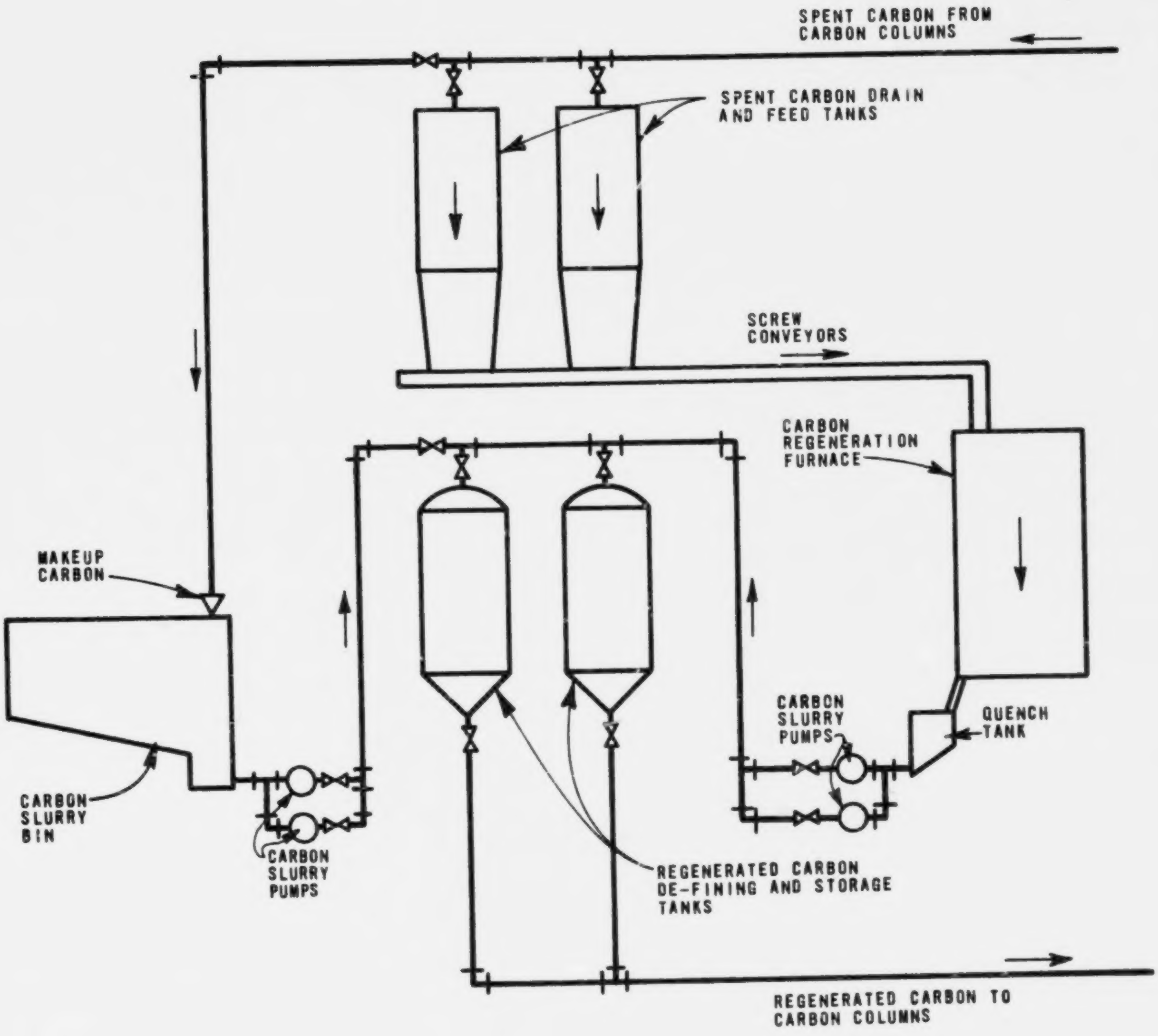
SPENT CARBON FROM CARBON COLUMNS
SPENT CARBON DRAIN AND FEED TANKS
SCREW CONVEYORS
CARBON REGENERATION FURNACE
MAKEUP CARBON
CARBON SLURRY PUMPS
QUENCH TANK
CARBON SLURRY BIN
CARBON SLURRY PUMPS
REGENERATED CARBON DE-FINING AND STORAGE TANKS
REGENERATED CARBON TO CARBON COLUMNS

FIGURE 9

CARBON REGENERATION SYSTEM

while spent carbon is replaced by fresh carbon. In addition, this procedure does not require firing the furnace every time carbon is withdrawn from the columns. To accomplish this, two concrete carbon storage bins have been provided, one to store spent carbon and the other to store regenerated and makeup virgin carbon.

Following the second regeneration of activated carbon, sand was added to each furnace hearth on the recommendation of the carbon manufacturer. The sand forms a layer between the hearth floor and the rabble arms on each hearth. A layer of sand is thus left on each hearth at the end of each regeneration, rather than a layer of carbon which would be burned to ash. Sand is used in this manner to reduce carbon losses in furnaces used for carbon production.

Carbon Regeneration Data

Time Required for Dewatering of Carbon. Samples of spent carbon were obtained from the carbon feed bins at a depth of 2 feet below the carbon surface when the feed to the furnace commenced. The time allowed for dewatering varied from 1 to 26 hours. The percent-moisture on a wet basis varied from 40.3 to 43.7 percent. In addition, samples were obtained at the point of discharge into the furnace and analyzed for moisture. The results fell within the range previously mentioned. It can be concluded that the dewatering of granular activated carbon can be obtained in 1 hour and that no apparent benefit can be realized at this time by retaining the carbon for an additional period of time.

Process Control Data. To date, approximately 148,500 pounds of carbon have been regenerated. Based upon the flow of 366 mg subjected to carbon treatment, the amount of carbon regenerated is equivalent to 405 pounds of carbon per million gallons of waste water treated. Current carbon dosages are less than 250 pounds/mg. The contact time in the main carbon columns has averaged 18 minutes.

Carbon Furnace Operating Data. The results of the carbon regenerations to date are shown in Tables 3 and 4. The furnace was initially fired to a temperature of 1,600°F, and carbon was fed at a rate of 240 pounds per hour (5,760 pounds per day). The burners on hearth 4 were set to deliver 4 percent excess oxygen and the burners on hearth 6, 2 percent excess oxygen.

When determining the benefit of adding steam to the furnace, the temperature was maintained at essentially 1,600°F and the feed at 240 pounds per hour.

Table 3

CARBON REGENERATION DATA

Date	Time	Item	Carbon Feed (lb./hr.)(dry)	BTU/Lb. Carbon	Apparent Density	Iodine Number	% Ash	Furnace Hearth Temperatures - °F					
								1	2	3	4	5	6
9/16/65	1500	Start Feed 1st Withdrawal Col. 1 — No Steam											
		Spent Carbon			0.536	715	4.46						
9/17/65	1000	Regenerated Composite	240	3,800	0.504	887	4.40	580	750	1,090	1,590	1,500	1,600
9/17/65	1030	Start Steam											
9/17/65	1400	Regenerated Composite	240	3,800	0.489	911	4.89	780	960	1,270	1,550	1,480	1,590
9/19/65	1830	Start Feed 2nd Withdrawal Col. 1 — With Steam											
		Spent Carbon			0.533	681	4.56						
9/20/65	0200	Regenerated Grab	400	2,250	0.504	880		710	880	1,170	1,640	1,520	1,640
9/20/65	0900	Regenerated Grab	240	3,800	0.497	935		810	1,020	1,240	1,580	1,460	1,650
9/20/65	0900	Start Feed Partially Spent Carbon											
		Spent Carbon			0.518	825							
9/21/65	0100	Regenerated Composite	235	3,920	0.491	915		750	950	1,300	1,650	1,550	1,650
9/21/65	0100	Start Feed 3rd Withdrawal Col. 1											
		Spent Carbon			0.530	806							
9/21/65	0800	Regenerated Composite	210	4,400	0.490	935		720	900	1,250	1,640	1,540	1,650
9/22/65	0100	Regenerated Composite	210	4,400	0.490	915							
9/22/65	0100	Start Feed 4th Withdrawal Col. 1											
		Spent Carbon			0.527								
9/23/65	0300	Regenerated Composite	200	(4,500)	0.485	900		840	1,020	1,250	1,640	1,540	1,640
9/23/65	0300	Start Feed 1st Withdrawal Col. 2											
		Spent Carbon			0.523	752							
9/24/65	0100	Regenerated Composite	240	(3,800)	0.486	954		820	990	1,300	1,640	1,540	1,670
9/24/65	0100	Start Feed 2nd Withdrawal Col. 2											
		Spent Carbon			0.525								
9/25/65	0100	Regenerated Composite	220	4,150	0.489			870	1,060	1,270	1,620	1,560	1,680
9/25/65	0100	Start Feed 3rd Withdrawal Col. 2											
		Spent Carbon			0.519								
9/25/65	1600	Regenerated Composite	240	3,800	0.473	909		850	1,030	1,320	1,620	1,550	1,670
10/13/65	2400	Start Feed, Bin 2											
		Spent Carbon	225	4,110		843							
		Regenerated Composite			0.495	938		1,000	1,200	1,380	1,640	1,540	1,660
10/15/65	0130	Start Feed, Bin 1											
		Spent Carbon	216	4,250		766							
		Regenerated Composite			0.488	922		880	1,080	1,380	1,660	1,580	1,700
10/16/67	0330	Start Feed, Bin 2											
		Spent Carbon	204	4,500		798							
		Regenerated Composite			0.485	902		810	990	1,220	1,670	1,590	1,670
10/17/65	0700	Start Feed, Bin 1											
		Spent Carbon	248	3,700	0.522								
		Regenerated Composite			0.493	962		750	920	1,260	1,640	1,560	1,670
10/18/65	0600	Start Feed, Bin 2											
		Spent Carbon	238	3,700									
		Regenerated Composite			0.483	969		570	680	960	1,630	1,560	1,660
10/19/65	0600	Start Feed, Bin 1											
		Spent Carbon	254	3,860									
		Regenerated Composite			0.480	969		740	910	1,240	1,620	1,530	1,640
10/19/65	1700	Stop 2nd Regeneration											

Table 4

CARBON REGENERATION DATA

Date	Item	Average Carbon Feed (lb./hr.) (dry)	BTU/Lb. Carbon	Apparent Density	Iodine Number	% Ash	Average Furnace Hearth Temperatures — °F					
							1	2	3	4	5	6
2/3/66	Bin 1											
	Regenerated	218	4,160	.498			830	1,040	1,350	1,660	1,580	1,690
2/4/66	Bin 2											
	Regenerated	220	3,920	.483			870	1,050	1,280	1,670	1,590	1,680
2/5/66	Bin 1											
	Regenerated	220	4,150	.483			800	990	1,310	1,650	1,560	1,660
8/23/66	Bin 2											
	Spent	258	2,680	.563	709	4.4	945	1,150	1,290	1,550	1,400	1,590
	Regenerated			.472	902	4.8						
8/24/66	Bin 1											
	Spent	254	3,990	.532	851	5.7	840	1,035	1,310	1,560	1,450	1,580
	Regenerated			.488	933	4.9						
10/19/66	Bin 1											
	Spent	183	4,620	.585	551	5.4	870	1,080	1,380	1,640	1,540	1,640
	Regenerated			.493	867	5.3						
10/21/66	Bin 2											
	Spent	190	5,590	.576	614	5.3	990	1,190	1,410	1,640	1,540	1,670
	Regenerated			.498	915	5.4						
11/8/66	Bin 1											
	Regenerated	196	2,840	.486			660	900	1,290	1,670	1,600	1,690
11/10/66	Bin 2											
	Spent	194	3,520		669	4.8	680	850	1,090	1,670	1,630	1,700
	Regenerated			.483	913	5.1						
11/11/66	Bin 1											
	Spent	174	3,870		550	5.3	710	940	1,320	1,680	1,600	1,690
	Regenerated			.492	842	5.6						
1/11/67	Bin 1											
	Spent	196	4,200	.574	615	4.8	840	1,040	1,370	1,660	1,540	1,680
	Regenerated			.493	814	5.7						
1/13/67	Bin 2											
	Regenerated	201	4,080	.488			870	1,050	1,270	1,630	1,560	1,660
4/17/67	Bin 1											
	Spent	184	5,310	.581	565	5.4	760	970	1,310	1,660	1,550	1,680
	Regenerated			.498	782	5.3						
4/19/67	Bin 2											
	Spent	181	5,300	.569	573	5.1	990	1,200	1,440	1,660	1,530	1,680
	Regenerated			.488	828	5.8						
4/22/67	Bin 1											
	Spent	170	6,950	.589	598	5.1	910	1,140	1,420	1,620	1,490	1,670
	Regenerated			.490	814	5.9						

By inspecting Table 3, and comparing the results of 9/17/65 — 1000 and 9/17/65 — 1400, the following may be concluded:

1. At a given temperature and feed rate, steam addition markedly reduced the apparent density (0.504 to 0.489) and increased the iodine number (887 to 911).

2. Steam addition gave a more uniform distribution of furnace hearth temperatures. Steam addition increased the temperatures of hearths 1, 2, and 3 by 200°F.

It was decided at this time that steam was to be used throughout the first few regenerations to study the efficiency of regeneration by varying steam, feed rate, and temperature. During these first few regenerations, 1 pound of steam was added per pound of dry carbon.

The following conclusions can be made about the effect of steam on carbon regeneration:

1. Steam gives a more uniform distribution of temperatures throughout the furnace.

2. A satisfactory degree of carbon regeneration can be obtained by adding 1 pound of steam per pound of dry carbon and maintaining a temperature of 1,650—1,680°F. By operating at these temperatures, it is believed that burning of carbon and the subsequent carbon losses will be minimized.

Table 5 shows the properties of virgin, spent, first-cycle regenerated, and second-cycle regenerated carbon. Thermal regeneration of granular activated carbon to essentially virgin characteristics was achieved during the first cycle, as indicated by the apparent density and iodine number. No buildup of ash was observed during the first cycling of the carbon. During the second cycle, an increase of approximately 1 percent ash was apparent.

Table 5

PROPERTIES OF CARBON AT VARIOUS
POINTS IN THE REGENERATION PROCESS

	Virgin Carbon	Spent Carbon	Carbon After One Regeneration With Steam	Carbon After Two Regenerations With Steam
Apparent Density	0.481	0.519—.589	0.473—0.497	.472—.498
Iodine Number	935	550—843	900—969	782—933
Ash-Percent	5.02	4.48	4.69	4.8—5.9

Carbon Losses. The cost of makeup carbon may constitute over 50 percent of the total cost of carbon treatment.(15) It is apparent that minimizing carbon losses is essential in order to operate an economical system. Sources of carbon losses include:

1. Shipping and handling
2. Poor housekeeping
3. Attrition in transferring carbon to and from the columns
4. Coring carbon in the screw conveyor feeding the furnace
5. Burning carbon in the furnace during regeneration and during procedures for startup and shutdown.

The most accurate method of determining carbon losses will be the measurements over a long period of time of the quantities of makeup carbon on a weight basis to maintain the columns at the desired depth. Based upon this period of operation, the carbon loss has averaged less than 5 percent per regeneration. This loss is about the average experienced in the sugar industry and that found in pilot-scale regeneration of carbon used for treatment of sewage,(14) but is less than that found at Pomona.

Sieve analyses were performed on the activated carbon at various points in the carbon transfer system and the regeneration process, in order to evaluate any change in carbon particle size due to physical handling of the carbon. The results of the sieve analyses are shown in Table 6. A comparison of the data indicates that there was some loss of 8-mesh carbon when passed through the furnace (II vs. III). The loss of 8-mesh carbon is balanced by a corresponding increase in the 16-mesh carbon size. Washing removed all particles finer than 200-mesh.

Table 6

SIEVE ANALYSES OF CARBON AT VARIOUS POINTS IN REGENERATION SYSTEM

		Percent Retained on Equivalent Screen Size							
Item	Sampling Point	4	8	16	30	50	100	200	<200
I	Spent Carbon	0	9.2	63.1	25.4	2.0	0.1	0.1	0.1
II	Spent Carbon at Discharge into Furnace	0	10.5	60.2	25.9	2.7	0.2	0.2	0.3
III	Washed Regenerated Carbon	0	8.2	62.8	25.7	3.0	0.1	0.2	0
IV	Virgin Carbon	0	6.8	58.1	32.0	2.6	0.2	0.1	0.2

Total Cost of Carbon Treatment. The estimated cost of activated carbon treatment and the related pretreatment at the South Tahoe P. U. D. plant are summarized in Table 7. The capital cost of the carbon system is higher than necessary for a plant with 2.5-mgd capacity because of the oversized carbon furnace which was provided in the Tahoe plant for future expansion. The furnace provided has capacity to regenerate the carbon required at the present dose to treat 7.5 mgd. The capital cost of the building, site preparation, etc., attributable to the carbon system was estimated by proportioning the total cost of the tertiary treatment plant on the basis of percentage of total plant area required for the carbon system.

The carbon regeneration cost is based on the period of operation to date, excluding with respect to carbon dose the initial startup period when operational difficulties resulted in excess carbon transfer. The increased fuel usage per pound (Table 4) has resulted in a slight increase in regeneration cost per pound. However, the low carbon loss (less than 5 percent) and the decreasing carbon dose have had the net effect of reducing the cost of carbon treatment. A dosage of 300 lbs/mg was used for cost analysis, although it appears the future dosage may be lower yet.

The cost of full-scale treatment, \$38.55/mg, is less than that projected from pilot plant studies by Joyce, et al,(14) who estimated the cost of carbon treatment to be \$70 per million gallons for a 10-mgd plant and \$130 per million gallons for a 1-mgd plant. It is also less than the estimate of \$80 per million gallons in a 10-mgd plant, as prepared at the Pomona, California installation.(16)

As mentioned, the carbon regeneration system has sufficient capacity to operate in conjunction with a 7.5-mgd plant. Thus, the estimated cost of activated carbon treatment at South Tahoe for a 7.5-mgd plant is only \$18.00/mg for capital costs and \$11.75/mg for regeneration costs, or a total of \$29.75/mg, compared to \$38.55/mg for the 2.5-mgd plant. Of course, chemical and regeneration costs are largely dependent upon the quality of effluent desired or necessary. Minimum water quality standards for the South Tahoe plant were presented on page 28.

The cost of the activated carbon treatment at South Lake Tahoe constitues about 28 percent of the total cost of the tertiary treatment.

PILOT CARBON COLUMN STUDIES

Description of Upflow Pilot Carbon Columns

In order to study the efficiency of the main carbon columns and the carbon regeneration system, four pilot carbon columns (PCC) were constructed. These columns were designed to operate upflow in parallel with the main carbon columns and at the same surface-loading rate.

Table 7

ESTIMATED COST OF ACTIVATED CARBON TREATMENT AND RELATED PRETREATMENT AT THE 2.5-MGD SOUTH TAHOE P. U. D. WATER RECLAMATION PLANT

		Cost Dollars/mg
Activated Carbon Treatment		
Capital Cost		
Carbon Columns	$ 48,332	
Regeneration System	89,445	
Activated Carbon	35,154	
Piping and Valves	29,000	
Electrical and Controls	20,000	
Building, Site Preparation, Earthwork, Concrete, Roofing, Painting, Testing, Paving, Cleanup, Heating, Ventilating, Miscellaneous	107,208	
Total	$330,139	
Capital Cost Amortized at 4% over 20 years at 2.5 mgd		$ 26.80
Regeneration Cost (5,000 lbs/day)		
Labor — $3.75/hr with 24 hr/day	$ 90.00/day	
Power — 11 hp at energy charge of $0.008/kwh and demand charge of $2.30/kw	3.46/day	
Fuel — 240 gals, propane/day at $0.086/gal	20.60/day	
	$114.06/day, or $.0228/lb	
With 5% Loss, Carbon Cost of $0.325/lb, Makeup Carbon Cost = $0.0163/lb Regenerated		
With Carbon Dose of 300 lbs/mg Regenerated Cost (.0228 + .0163)(300) =		11.75
Subtotal (Activated Carbon Treatment Cost)		$ 38.55
Cost of Carbon Pretreatment (Separation Bed Process)		
Capital Cost		
$284,861 Amortized at 4% over 20 years at 2.5 mgd		$ 22.80
Chemicals		
Alum (Dose for Phosphate Removal) 200 mg/l at $57.40/ton		47.60
Polyelectrolyte — 0.8 mg/l at $1.25/lb		8.35
Labor		
6 hrs/day at $3.75/hr for 2.5 mg		9.00
Power		
Based on 10-minute Backwash Every 4 hrs at $0.008/kwh Energy Charge and $2.30/kw Demand Charge		10.50
Subtotal (Separation Bed Treatment)		$ 98.25
Subtotal (Activated Carbon Treatment)		38.55
Total Cost of Tertiary Treatment		$136.80

A 1-inch tap was installed in the influent header to the main carbon columns. The separation bed effluent is pumped upwards through the four pilot carbon columns. The rate of flow to each column is regulated by a rotameter, 0- to 55-gph capacity, and the total volume of separation bed effluent passing through the columns is measured by a water meter.

These pilot columns are constructed of 4-inch-diameter PVC pipe and have an effective depth of 14'-6", including pipe and couplings. (Figure 10) Six 1/4-inch sampling taps are provided to enable measurement of adsorbate concentrations at various carbon depths. A 4-inch-diameter Neva-clog screen holds the carbon in place at the bottom of the column. A 30-mesh stainless steel screen is located at the top of the column. Provisions are made at the top and bottom of the column for the addition of fresh carbon and the removal of spent carbon.

Description of Downflow Pilot Carbon Columns

To adequately study carbon adsorption, downflow pilot columns of the four-stage type were also operated. Two of these columns (Figures 11 and 12) were constructed of 6-inch-diamenter plexiglass in four, 5-foot sections. A 30-mesh stainless steel screen is located at the bottom of each section to hold the carbon in place. Sampling and pressure taps are provided after each section. Valves and piping are arranged so that any section can be the lead section and any can be backwashed as necessary.

The rate of flow through each column is controlled at the effluent by a 0- to 55-gph-capacity rotameter, andthe total volume of water passed is measured by a water meter.

Water from any stage of the treatment can be passed through these columns as desired.

Experimental Procedures

Run 1. The initial head loss in the main carbon columns was considerably higher than anticipated. Because the initial carbon charge to the main columns was not washed in order to remove carbon fines which might have been generated during shipping, it was possible that excessive amounts of carbon fines were contributing to the high head loss in the main columns. One pilot carbon column (PCC 4) was charged with washed 8 x 30-mesh carbon to determine if carbon fines, which could be removed by washing, were contributing to the high head loss in the main columns.

Two pilot carbon columns (PCC 1 and 2) were charged with unwashed 8 x 30-mesh carbon, as were the main columns. These two pilot columns were to be operated at the same surface loading rate as the main columns. However,

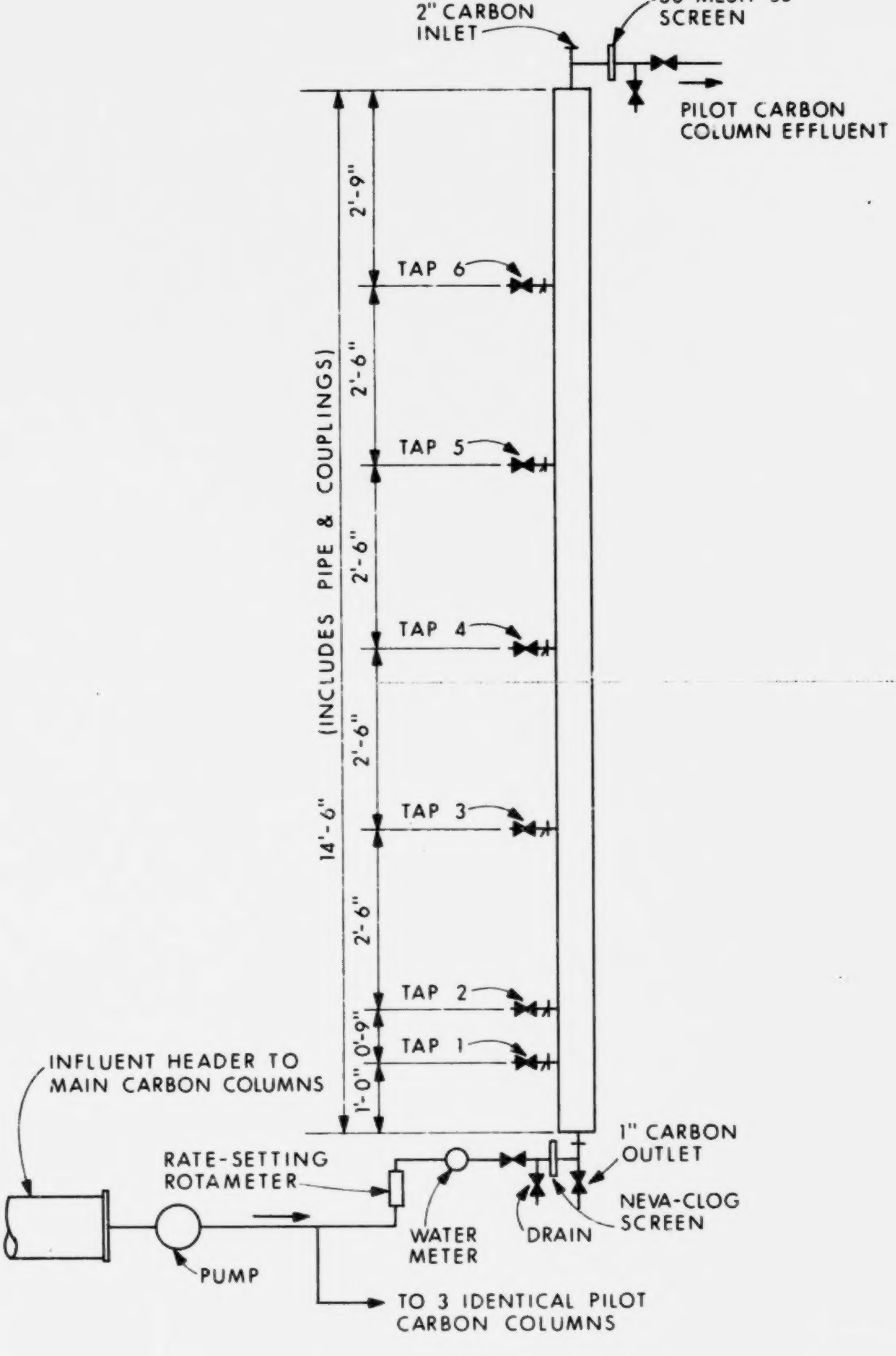
2" CARBON INLET
30-MESH SS SCREEN
PILOT CARBON COLUMN EFFLUENT
2'-9"
TAP 6
2'-6"
TAP 5
2'-6"
TAP 4
2'-6"
TAP 3
(INCLUDES PIPE & COUPLINGS)
14'-6"
2'-6"
TAP 2
0'-9"
TAP 1
1'-0"
INFLUENT HEADER TO MAIN CARBON COLUMNS
1" CARBON OUTLET
RATE-SETTING ROTAMETER
NEVA-CLOG SCREEN
WATER METER
DRAIN
PUMP
TO 3 IDENTICAL PILOT CARBON COLUMNS

FIGURE 10

UPFLOW PILOT CARBON COLUMNS

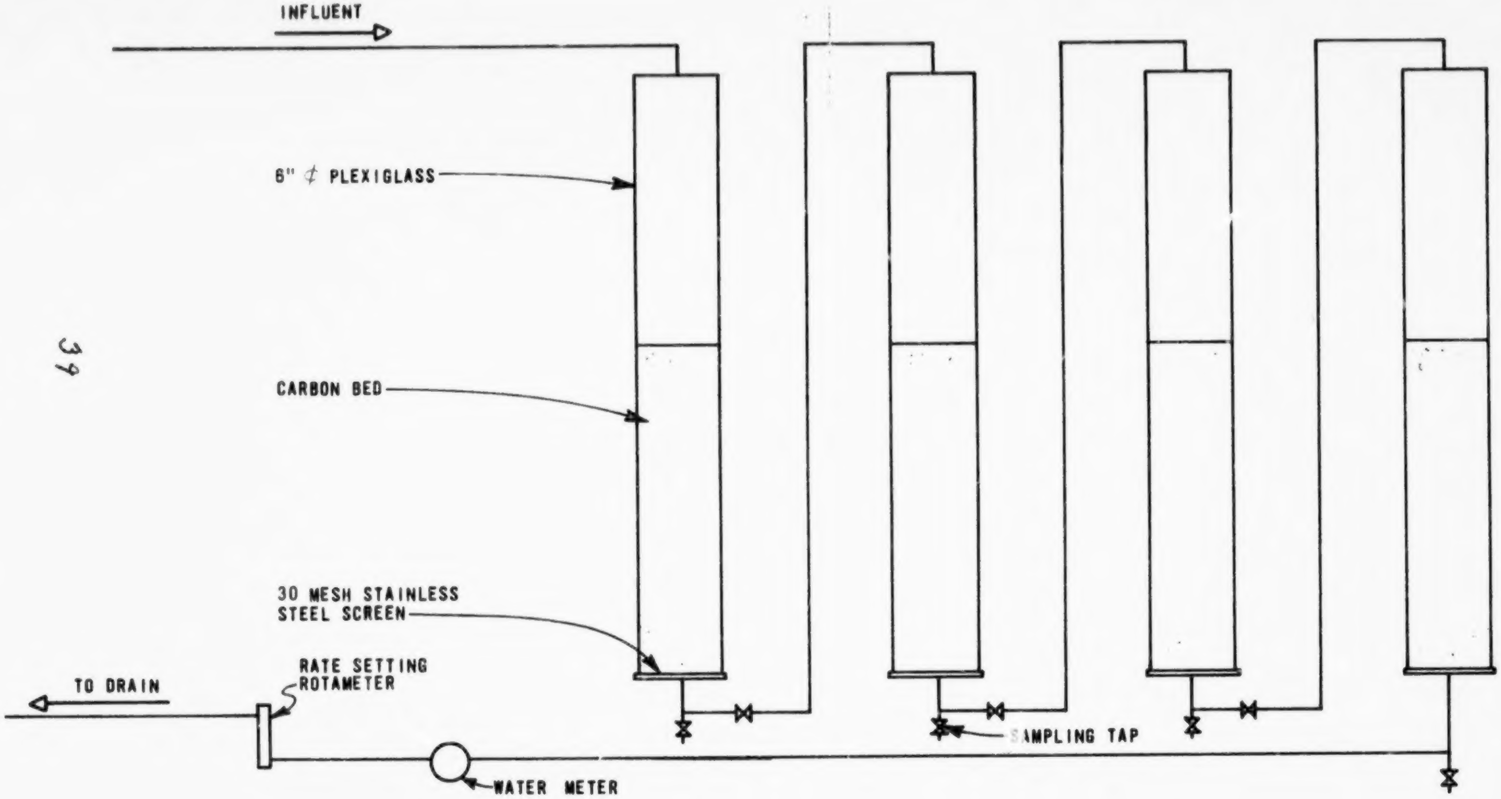
INFLUENT
6" ⊄ PLEXIGLASS
CARBON BED
30 MESH STAINLESS STEEL SCREEN
RATE SETTING ROTAMETER
TO DRAIN
WATER METER
SAMPLING TAP

FIGURE 11
DOWNFLOW PILOT CARBON COLUMNS

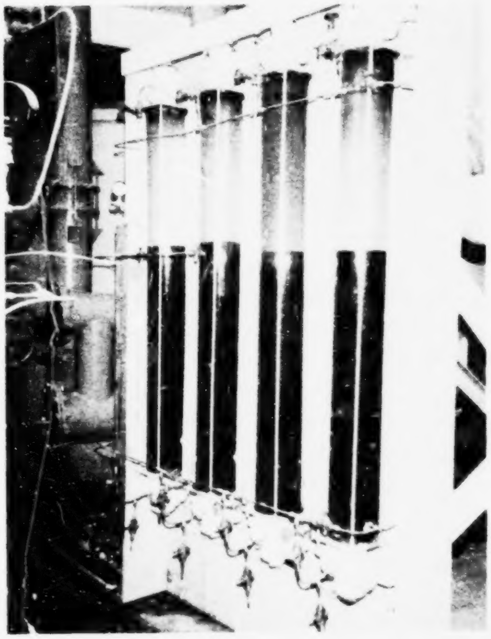

FIGURE 12
DOWNFLOW PILOT CARBON COLUMNS

excessive head loss across these pilot columns reduced the flows to such an extent that their loading was no longer in parallel with that being applied to the main columns. The flow through PCC 4, which contained washed 8 x 30-mesh carbon, was maintained at a rate comparable to that in the main columns. Thus, the data from PCC 1 and 2 will be compared with that from PCC 4 to determine the effects of the different loading rates on adsorption.

The remaining pilot column (PCC 3) was charged with unwashed 12 x 40-mesh carbon to measure the effect of carbon particle size on column performance. Three days after the pilot carbon effluent screens were changed, the flow through the 12 x 40-mesh carbon fell off. Inspection of the effluent screen showed that no fines were plugging the screen. The carbon was then removed from the column so that the column could be charged with washed 12 x 40-mesh carbon. It was found that the lower portion of the column was plugged with a fine floc. It is believed that this entrapped floc, as well as carbon fines, were contributing to the excessive head loss across the pilot columns and the main columns.

To alleviate high head loss problem, about 10 percent of the original carbon charge was removed from the main carbon columns and pilot carbon columns to obtain some freeboard for washing the carbon column and to allow room for expansion of the carbon during normal operation. It was found that the excessive amounts of head loss across the columns could be eliminated by occasionally downflushing the columns. Downflushing not only removed the carbon fines from the effluent screens, but washed the floc from the lower portion of the column. A considerable amount of floc was observed to be present in the backflush water emerging from the pilot carbon columns during the initial period of backflushing.

Run 2. One pilot carbon column (PCC 5) was charged with virgin 8 x 30-mesh carbon and operated in parallel with another column (PCC 6), which was charged with regenerated 8 x 30-mesh carbon. The regenerated carbon used in PCC 6 was a composite sample of the regenerated carbon from the first carbon regeneration.

Pilot carbon column 4, which contained the washed 8 x 30-mesh carbon used in run 1, was operated in parallel with the main columns. Spent carbon was withdrawn and regenerated carbon added in the same proportions as in the main columns.

Run 3. Again, the four upflow pilot columns were charged with various types of carbon to determine the relative efficiencies for the several carbons. PCC 12 contained the 8 x 30-mesh Pittsburgh type SGL carbon used as a control; PCC 13 was charged with 8 x 30-mesh Brand A carbon; PCC 14 contained 12 x 30-mesh Brand B, activated charcoal; PCC 14 was charged 8 x 40-mesh Brand C carbon.

Run 4. Downflow pilot columns (PCC 10 and PCC 11) were each charged with 60 lbs of 8 x 30-mesh Pittsburgh type SGL carbon (15 lbs in each stage). PCC 10 was operated treating secondary effluent, while PCC 11 had separation bed effluent passed through it. The total contact time was 40 minutes for both columns.

The purpose of this study was to determine the effects of pretreatment on carbon column operation, as well as the results of longer contact times.

Sampling and Analysis. Composite samples of the separation bed effluent (carbon column influent) were collected daily. Grab samples were obtained from all the taps on a particular column every other day. In addition, a grab sample of the carbon column influents was obtained at the same time the pilot column taps and main carbon columns were sampled.

All of these samples were analyzed initially for apparent alkyl benzene sulfonate (ABS), chemical oxygen demand (COD), and total organic carbon (TOC). Analysis for ABS and COD were performed by project personnel, while the TOC analyses were performed at the Robert A. Taft Sanitary Engineering Center, Cincinnati, Ohio.

During the course of the studies, the increased use of biodegradable detergents resulted in a marked decrease in the apparent ABS concentration in the waste water. Analysis for surfactants was therefore discontinued on the pilot studies following Run 2. At the same time, TOC data collection on the pilot column work was discontinued due to the heavy workload on the Cincinnati laboratory. Both ABS and TOC analyses have been continued for the main plant, however.

Experimental Results of Runs 1 and 2

The operational data obtained from the pilot carbon columns and main carbon columns are shown in Table 8 and as breakthrough curves (Figures 13—22). Some of the COD and TOC sampling depths have been eliminated for the sake of clarity.

As previously mentioned, PCC1 and 2 operated at a lower surface-loading rate than the main carbon columns. Since PCC1 and 2 were operated at essentially the same surface-loading rate, only the results from PCC 2 will be presented. Data analysis showed no significant difference in the results obtained from PCC 1 and 2.

Pilot carbon columns 5 and 6, charged with virgin and one cycle regenerated 8 x 30-mesh carbon, respectively, were operated in parallel. The operational data on PCC 5 and 6 are shown on Figures 19—22 as concentration curves for ABS and TOC.

Table 8

CARBON COLUMN OPERATING DATA – RUN ONE

Column	Carbon[1]	Average Surface Loading Rate (gpm/ft^2)	Average Residence Time (Minutes)	Volume Passed (gallons)	Bed Volumes Passed	Applied Loading			Average Influent Conc. (mg/l)			Effluent Conc. at End of Run (mg/l)		
						Lbs ABS	Lbs COD	Lbs TOC	ABS	COD	TOC	ABS	COD	TOC
PCC 1	Virgin 8 x 30	4.5	24.2	12,752	1,372	0.142	3.58	1.49	1.34	33.6	11.7	.002	2.5	2.0
PCC 2	Virgin 8 x 30	4.1	26.4	9,967	1,070	0.137	3.18	1.25	1.65	38.3	11.0	.002	1.5	1.0
PCC 3	Washed Virgin 12 x 40	6.5	16.7	13,138	1,415	0.158	3.56	1.23	1.44	32.7	10.7	.003	10	3.0
PCC 4	Washed Virgin 8 x 30	6.5	16.7	18,226	1,970	0.218	5.45	1.80	1.43	35.9	11.0	.12	5	4.0
Main Columns	Virgin 8 x 30	7.6	13.7	55.3×10^6	2,040	759.5	16,055	5,797	1.65	35.0	12.6	0.5	26	7.0

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(1) All carbon is Pittsburgh Activated Carbon Co., Type SGL.

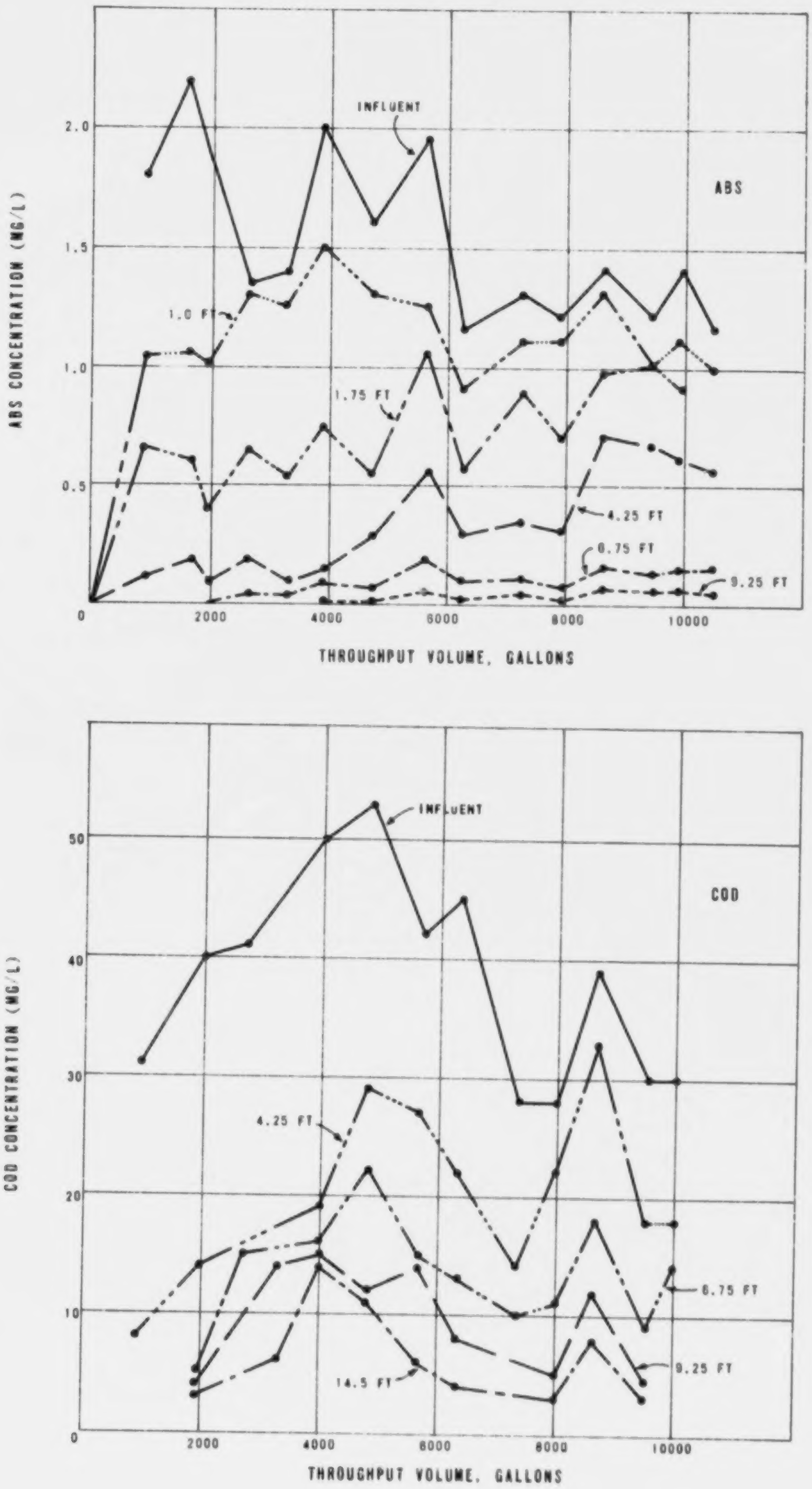


FIGURE 13
ABS AND COD BREAKTHROUGH CURVES FOR PCC 2
(8X30 MESH CARBON - 4.1 GPM/FT^2)

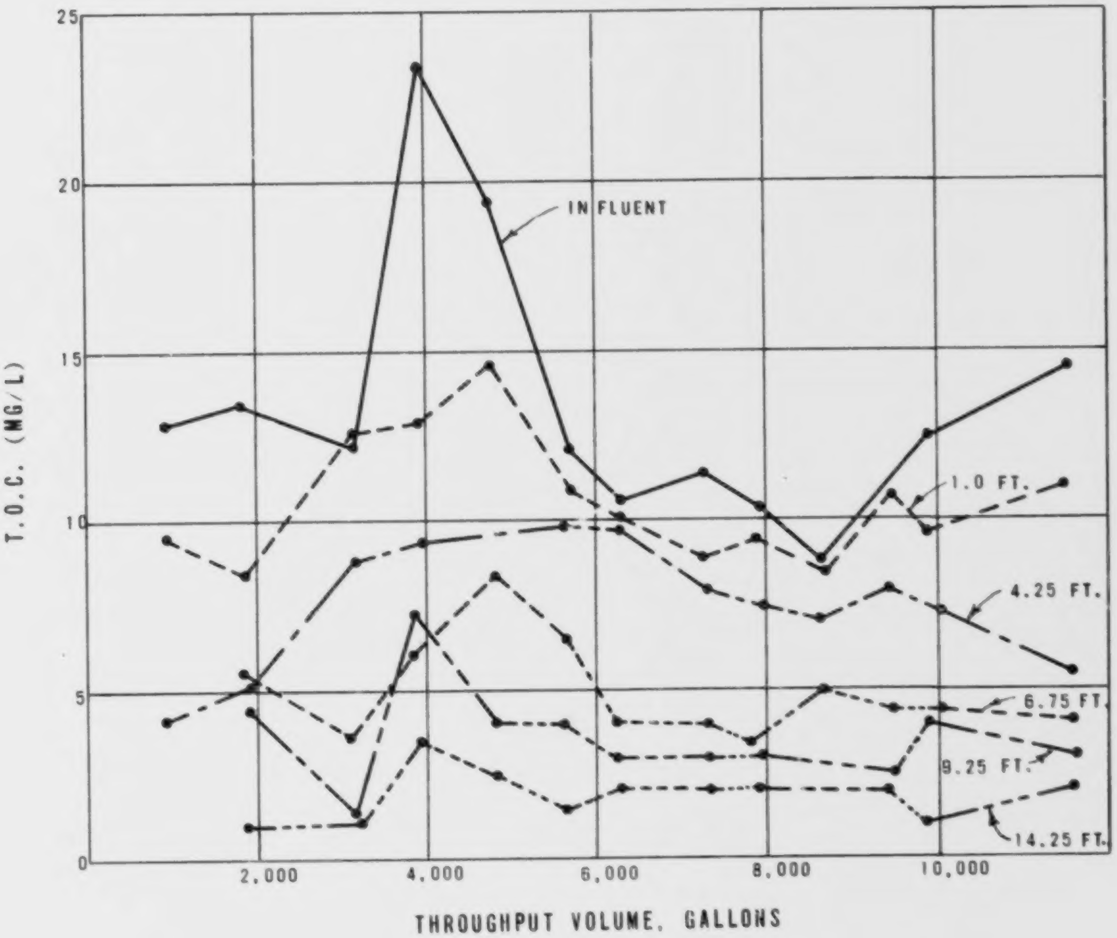


FIGURE 14
T.O.C. (TOTAL ORGANIC CARBON) BREAKTHROUGH CURVE FOR PCC 2
(8X30 MESH CARBON - 4.1 GPM/FT^2)

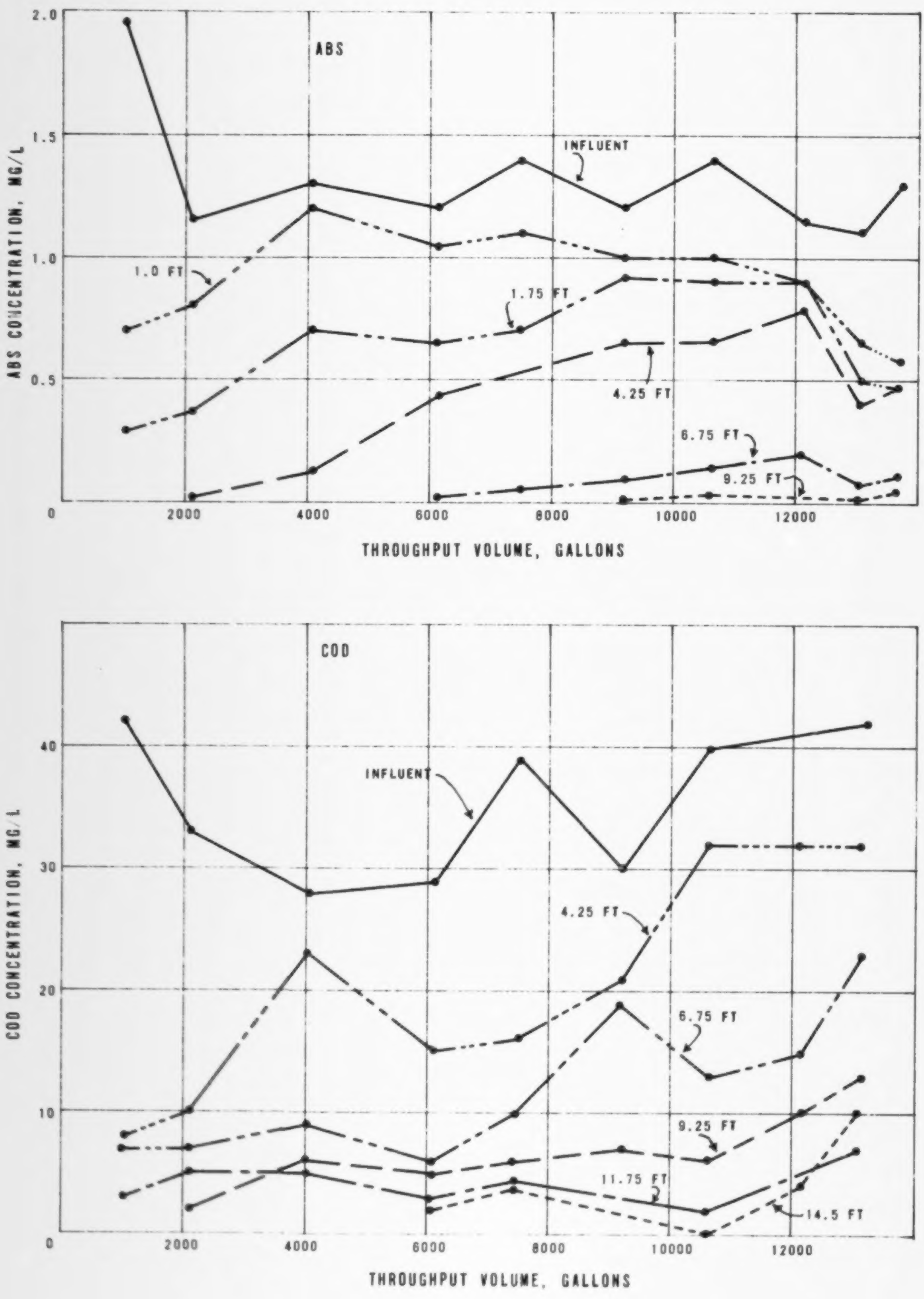


FIGURE 15

ABS AND COD BREAKTHROUGH CURVES FOR PCC 3

(12X40 MESH CARBON - 6.5 GPM/FT^2)

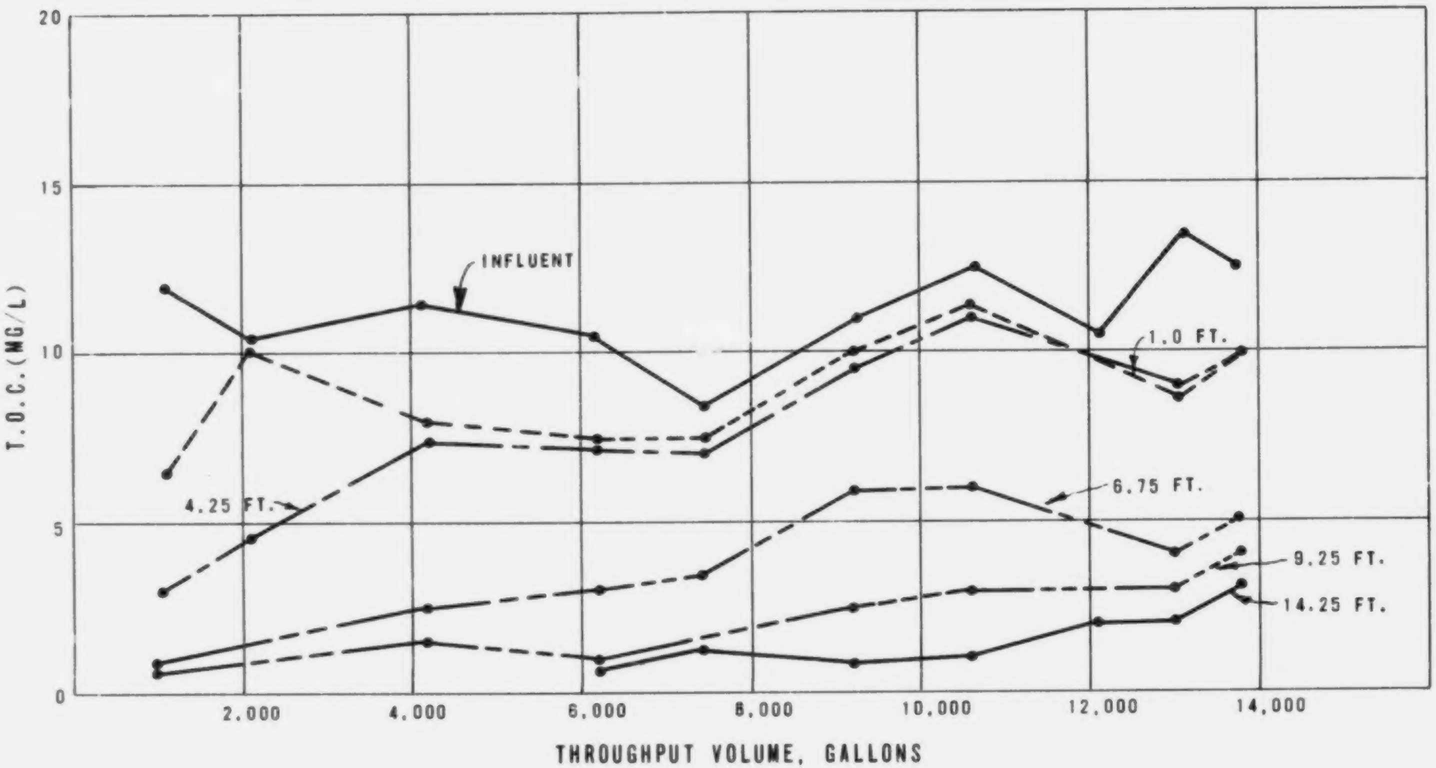
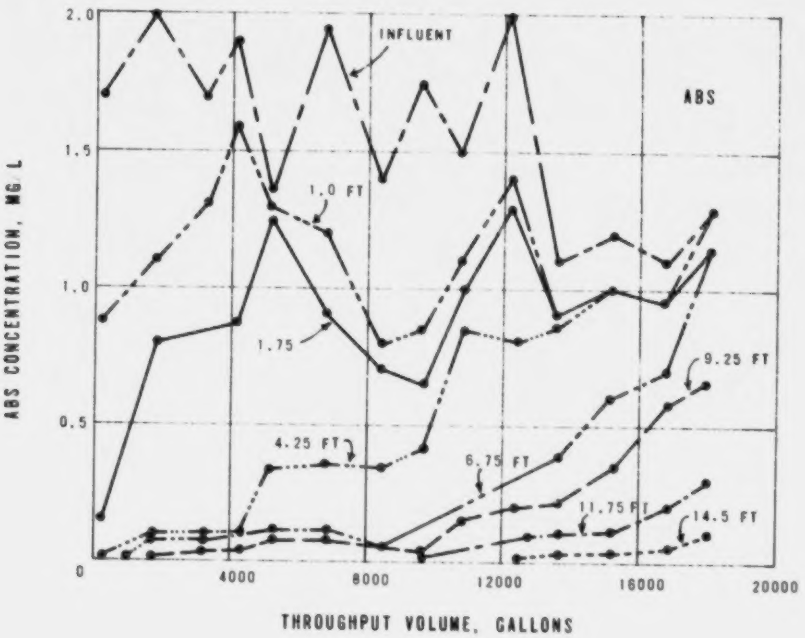


FIGURE 16
T.O.C. (TOTAL ORGANIC CARBON) BREAKTHROUGH CURVE FOR PCC 3
(12X40 MESH CARBON - 6.5 GPM/FT^2)



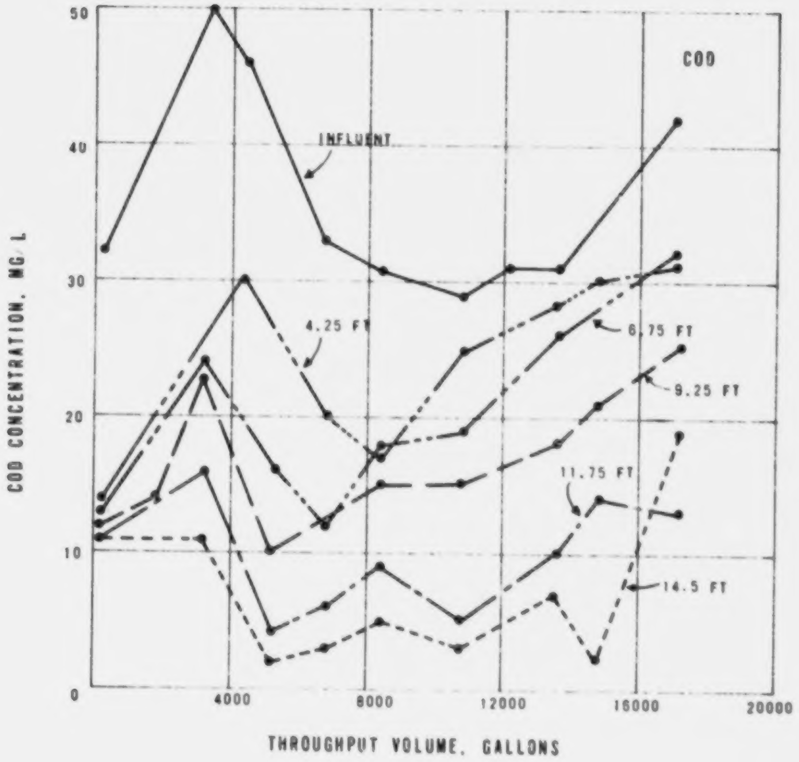


FIGURE 17

ABS AND COD BREAKTHROUGH CURVES FOR PCC 4
(8X30 MESH CARBON - 5.5 GPM/FT^2)

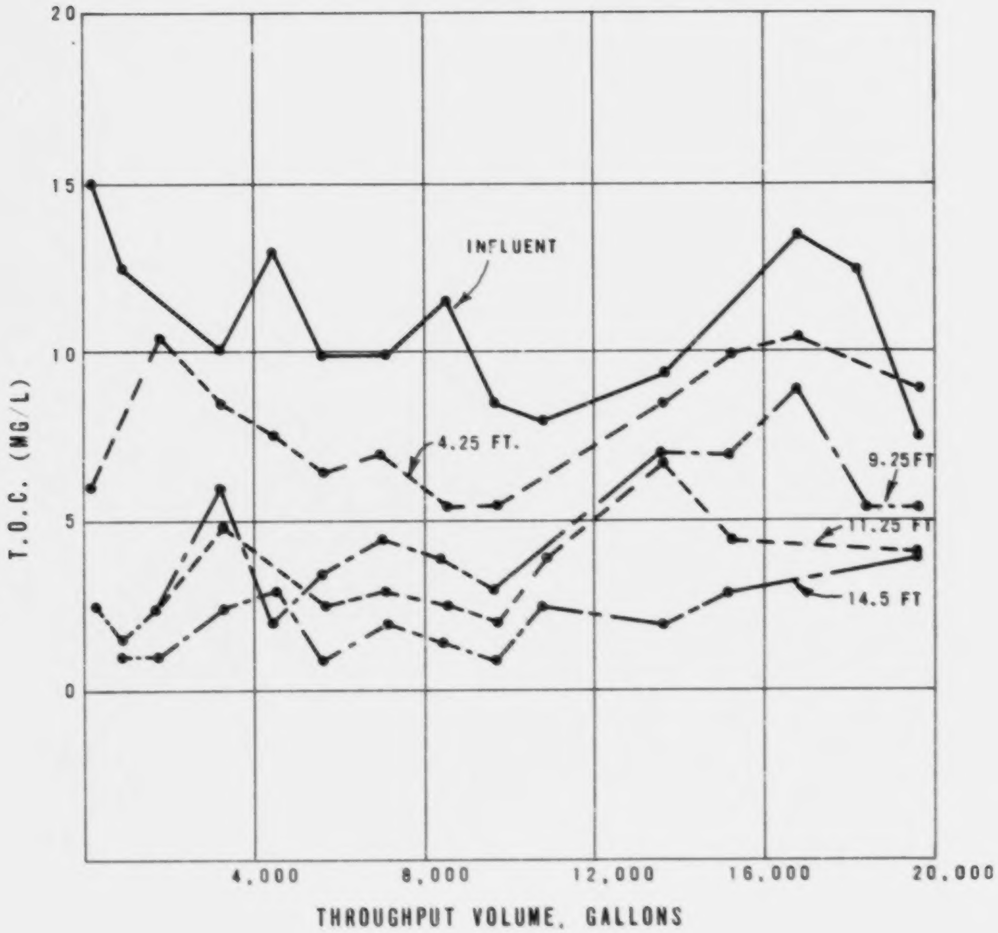


FIGURE 18
T.O.C. (TOTAL ORGANIC CARBON) BREAKTHROUGH CURVE FOR PCC 4
(8X30 MESH - 6.5 GPM/FT^2)

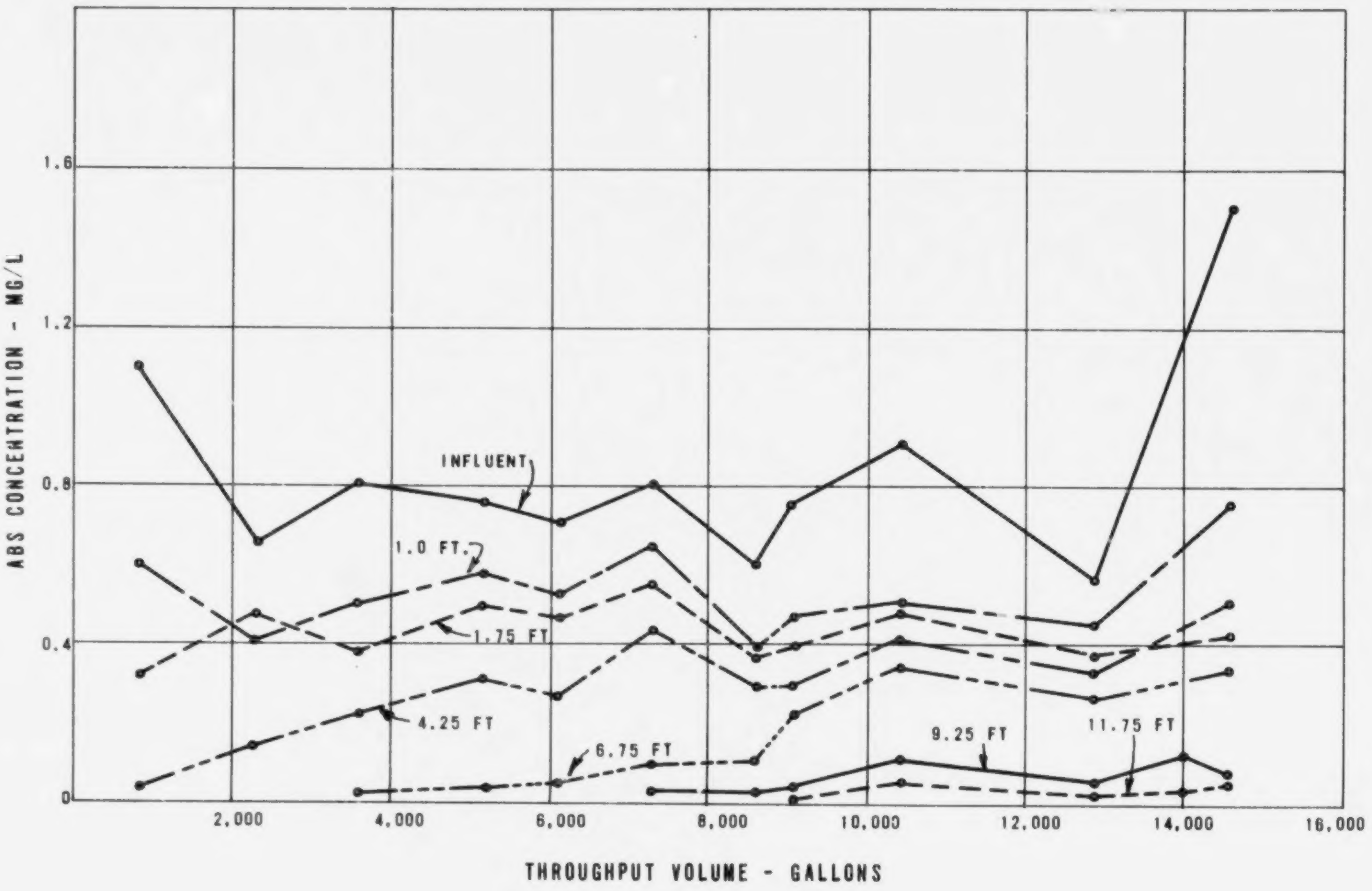


FIGURE 19

ABS BREAKTHROUGH CURVE

PCC 5

(30 LBS. 8X30 MESH VIRGIN CARBON - 4.6 GPM/FT^2)

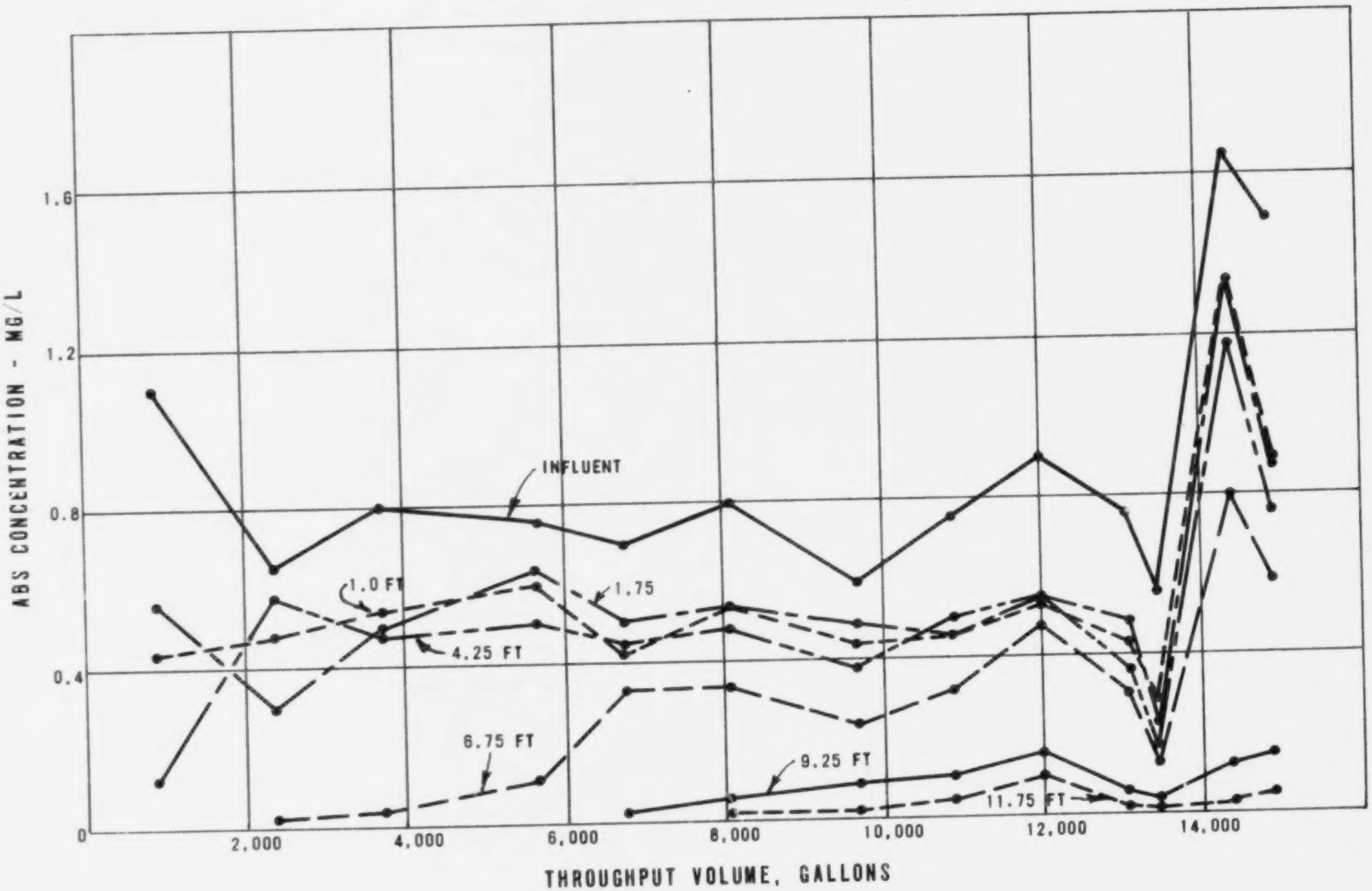


FIGURE 20

ABS BREAKTHROUGH CURVE

PCC 6

(30 LBS. 8X30 MESH 1ST CYCLE REGENERATED CARBON - 4.6 GPM/FT^2)

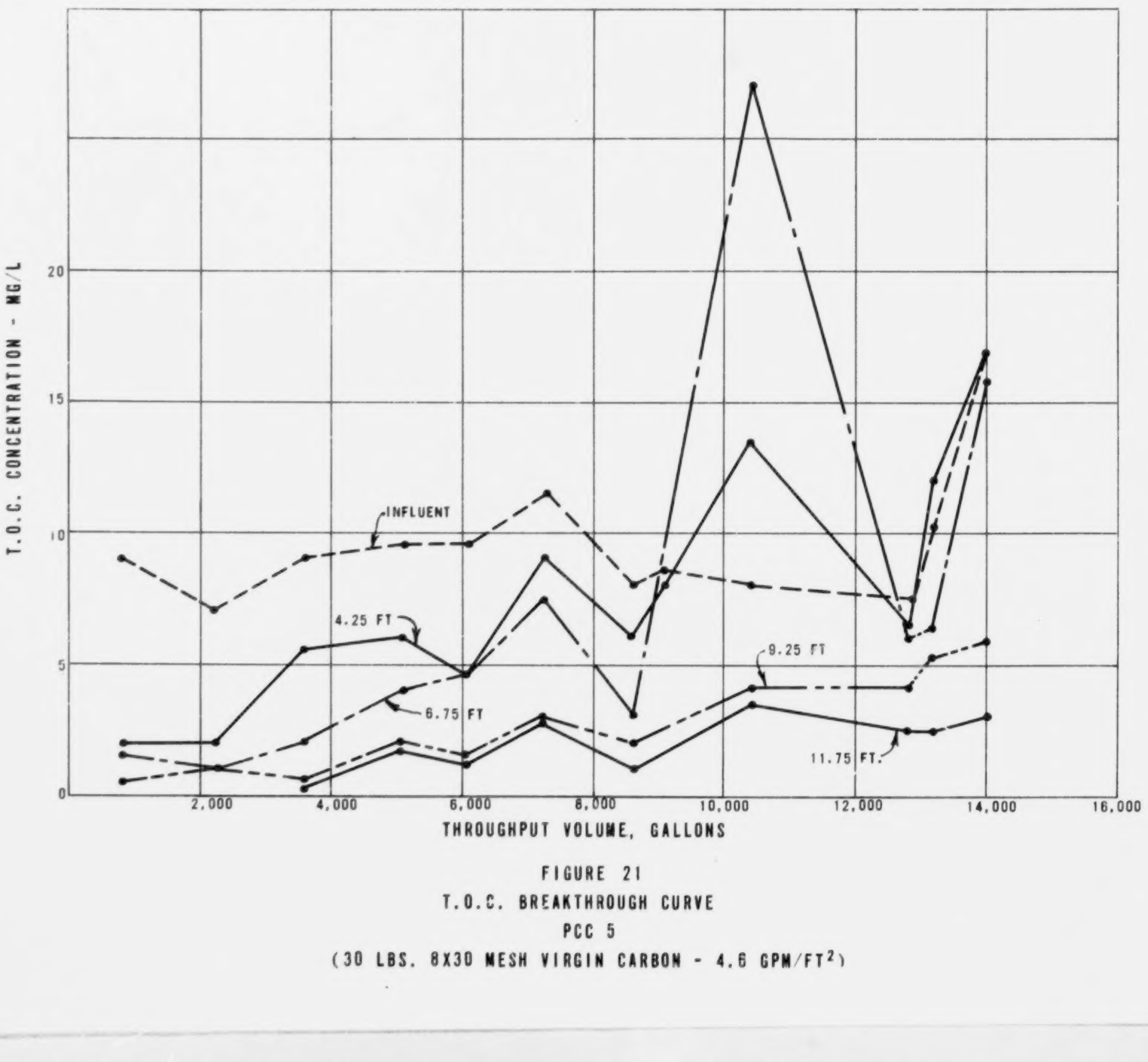


FIGURE 21
T.O.C. BREAKTHROUGH CURVE
PCC 5
(30 LBS. 8X30 MESH VIRGIN CARBON - 4.6 GPM/FT^2)

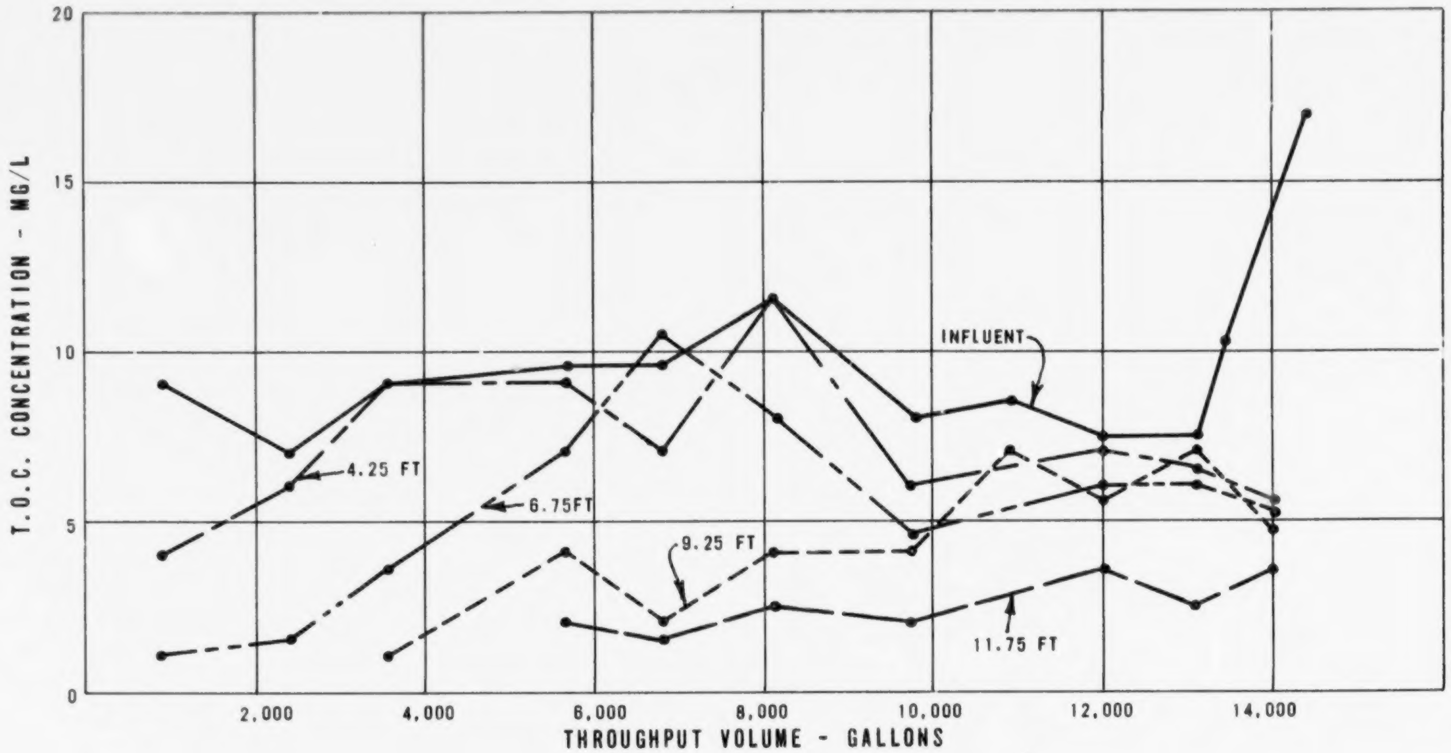


FIGURE 22

T.O.C. BREAKTHROUGH CURVE

PCC 6

(30 LBS. 8X30 MESH 1ST CYCLE REGENERATED CARBON - 4.6 GPM/FT^2)

Pilot carbon column 4 was to be operated in parallel with the main carbon columns, but mechanical problems with this pilot column eventually made it impossible to maintain a rate of flow parallel to that in the main carbon columns. The use of the pilot column had to be discontinued.

Effect of Mechanical Fouling of Carbon. As previously discussed, a poor quality secondary effluent was being treated during the first run. As a result, the turbidity of the water applied to the carbon columns was higher than desired. The plugging of the columns with this turbidity (often more than one unit) was reflected by an increasing head loss across the columns with continued operation. This head loss could be reduced by backflushing the carbon columns to flush the entrapped floc from the bottom of the columns. In addition to increasing the head loss across the carbon columns, the turbidity trapped in the lower portions of the columns resulted in mechanical fouling of the carbon.

Figure 23 shows that the first foot of carbon continued to adsorb ABS and COD at a more rapid rate than the increment of carbon immediately above the first foot of carbon which had essentially ceased to remove adsorbate, even though this increment had not adsorbed as much material as the lower foot of carbon. This decrease in adsorptive capacity is believed to have been caused by mechanical fouling of the carbon. There are several points on these curves where the rate of adsorption suddently rises in the 0—1.0-foot increment after having reached a plateau. These sudden increases in the rate of adsorption occurred at times when the column was downflushed. This increase in the rate of adsorption and the previously mentioned observations can be attributed to the fact that the carbon was mechanically fouled and that by downflushing the carbon, the floc particles coating the carbon granules were removed from a portion of the carbon and that adsorption was again possible. As Figures 23 and 24 show, downflushing did not remove the floc at a depth of 1.75 to 4.25 feet. The floc particles apparently had not fouled the carbon significantly beyond 4.25 feet, as the carbon continued to adsorb beyond this point.

Figure 25 shows the ABS profile for PCC 4 after various increments of flow. These profiles also reflect the effect of mechanical fouling. Although the lowest portion of the column continued to reduce the influent concentration, Co, the concentration, after 13,560 gallons had passed, was not reduced further by the next segment of carbon. The ABS concentration was then reduced as the water reached the carbon which was not fouled.

The principle underlying the countercurrent operation of the carbon columns is clearly demonstrated by the data shown on Figures 23 and 24 and Table 9. It is evident that the bulk of the adsorption is taking place in the first foot of the carbon column. By operating the column in a countercurrent manner, only the most saturated carbon will be withdrawn for regeneration.

Based upon data from PCC 4, the bottom one foot of carbon had adsorbed 2.8 percent ABS by weight and 45 percent COD by weight prior to the first

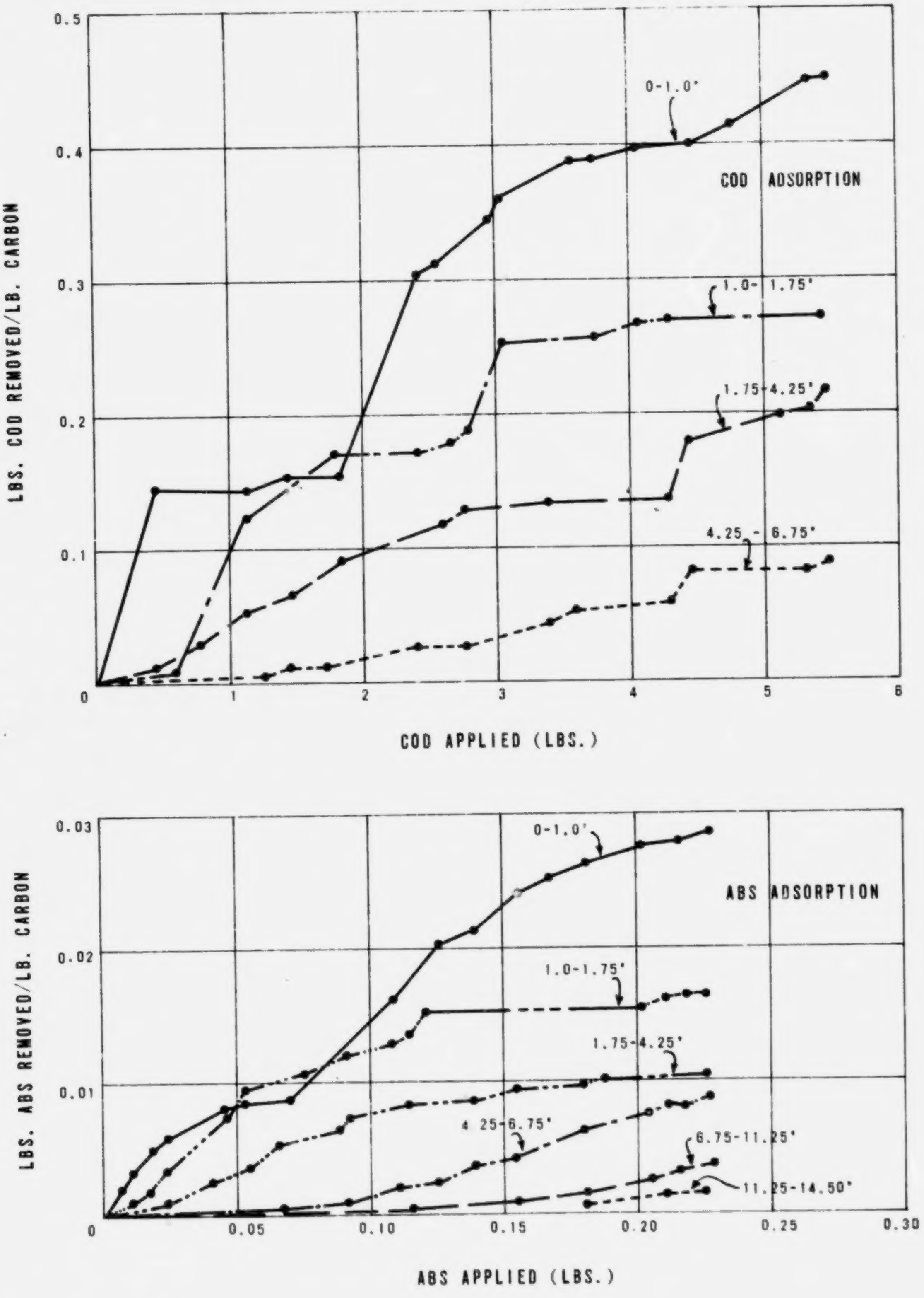


FIGURE 23

COD AND ABS ADSORPTION PILOT CARBON COLUMN 4

(VIRGIN 8X30 MESH CARBON

6.5 GPM/FT^2)

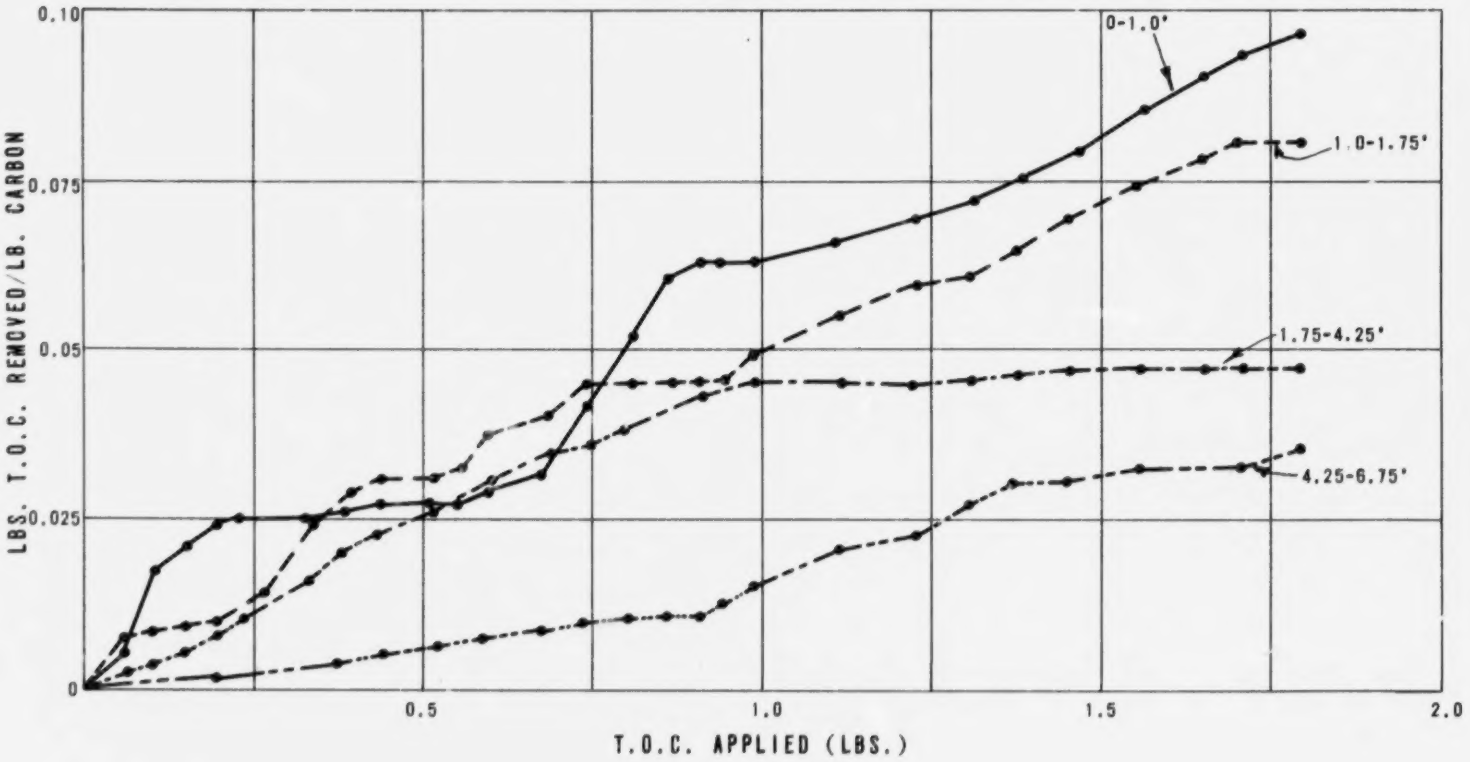


FIGURE 24
T.O.C. ADSORPTION
PCC 4

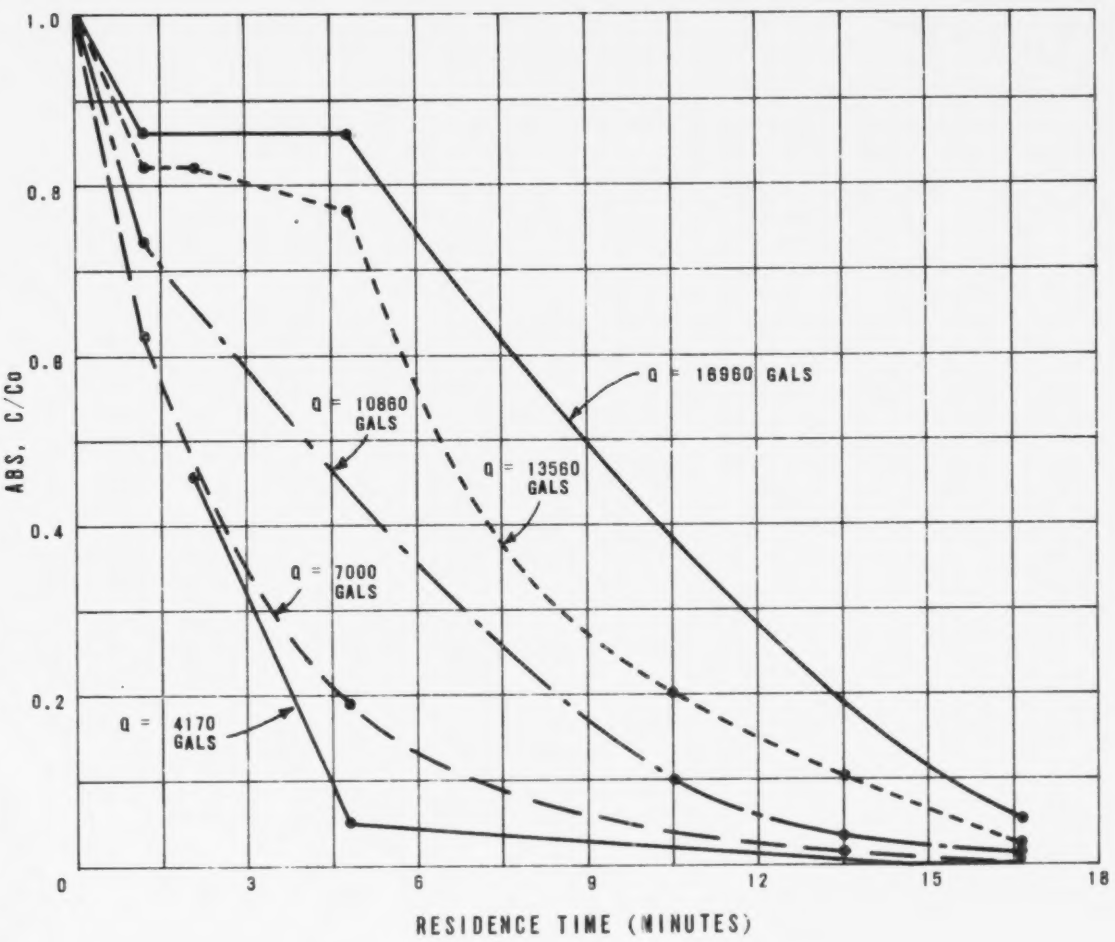


FIGURE 25

ABS ADSORPTION PROFILE FOR VARIOUS FLOWS, PCC 4

VIRGIN 8X30 MESH CARBON, 6.5 GPM/FT^2

withdrawal. Due to the mechanical fouling discussed earlier, carbon at higher levels in the column was no longer adsorbing organic materials, although it was not saturated to the same degree as the bottom one foot of carbon. Because of this fouling, the carbon withdrawn for the first regeneration, although no longer adsorbing at a significant rate, averaged only 0.9 percent ABS adsorption by weight and 17 percent COD adsorption by weight.

Table 9

PERCENT SATURATION OF CARBON FOR INCREMENTS OF DEPTH AT END OF RUN 1 FOR PCC 4

				Percent Saturation Relative to First Foot of Carbon		
Depth Feet	Lbs COD Removed Lbs Carbon	Lbs ABS Removed Lbs Carbon	Lbs TOC Removed Lbs Carbon	COD	ABS	TOC
0−1.0	0.448	.0285	.096	100	100	100
1.0−1.75	0.270	.0163	.082	60.3	57.4	85.5
1.75−4.25	0.218	.0102	.047	48.6	35.9	49.0
4.25−6.75	0.085	.0085	.034	19	29.8	35.0
6.75−11.25	−	.0035	−	−	12.3	−
11.25−14.50	−	.0015	−	−	5.4	−

Table 10 shows the quantities of ABS, COD, and TOC removed by the pilot carbon columns. The carbon columns removed 98 percent of the applied ABS and approximately 80 percent of the applied organic loading.

The carbon dose required to obtain the minimum effluent water quality standards is dependent not only on the efficiency of the carbon but on the volumetric efficiency of the carbon columns. The volumetric efficiency is dependent upon a uniform distribution of flow throughout the carbon bed and a uniform withdrawal of spent carbon from the column.

To determine if a uniform distribution of flow was being obtained, a comparison of the concentration curves for the main columns and PCC 4 was made. A comparison of the effluent quality at a residence time of 13.5 minutes for the main carbon columns and PCC 4 is shown on Figure 26 and Table 11. The correlation between the two columns is within experimental error, and it can be concluded that a uniform distribution of flow is being achieved in the main carbon columns.

It was intended to withdraw carbon from PCC 4 in direct proportion to that withdrawn from the main carbon columns and to continue operating PCC 4 in parallel to the main carbon columns in order to determine if a uniform withdrawal

Table 10

QUANTITIES OF ABS, COD, TOC REMOVED

Pilot Carbon Column	ABS			COD			TOC		
	Lbs Applied	Lbs Removed	Percent Removed	Lbs Applied	Lbs Removed	Percent Removed	Lbs Applied	Lbs Removed	Percent Removed
1	.1517	.1494	98.5	3.575	3.078	86	1.4920	1.2145	81.5
2	.1560	.1521	98	3.1750	2.738	86	1.2516	1.0597	84.5
3	.1643	.1632	99.5	3.558	2.843	80	1.2313	0.9824	79.5
4	.2264	.2224	98.5	5.446	4.186	77	1.7989	1.3636	76.0

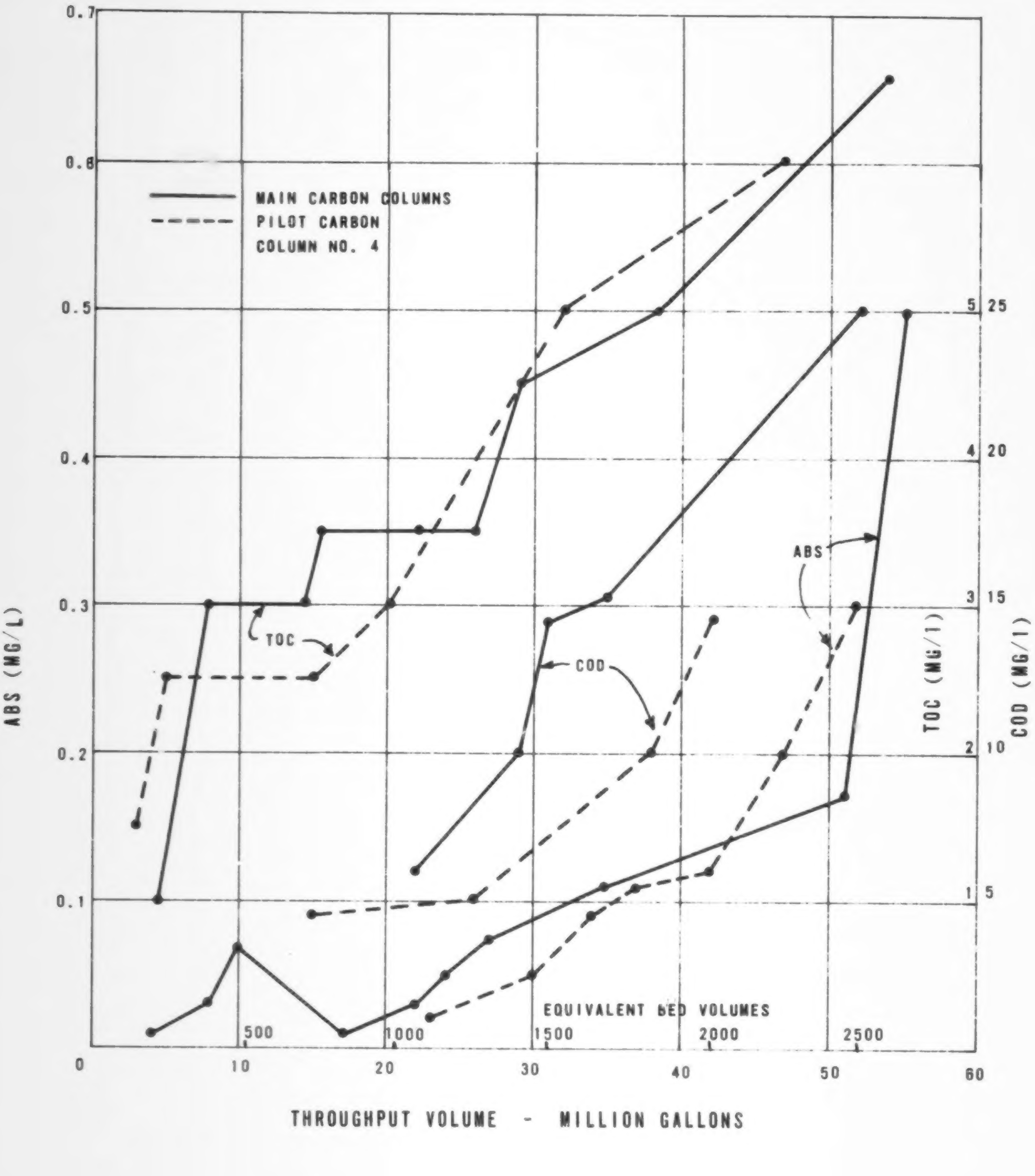


FIGURE 26

COMPARISON OF EFFLUENT QUALITY
AT 13.5 MINUTE RESIDENCE TIME
PCC 4 AND MAIN CARBON COLUMNS

of spent carbon is being obtained in the main carbon columns. However, the head loss across PCC 4 became so excessive that it was impossible to pass a significant volume of water through this column. Attempts to clean PCC 4 by backflushing were futile, and this use of the column was abandoned. The volumetric efficiency of the main carbon columns will now be determined by comparing the decrease in efficiency of the regenerated carbon and the corresponding increase in carbon dose in the main columns.

Table 11

COMPARISON OF EFFLUENT QUALITY
AT 13.5 MINUTES RESIDENCE TIME
AFTER 2,000 BED VOLUMES

	Full-Scale Carbon Columns(1)	Pilot Carbon Column No. 4(1)
TOC, mg/l	5.4	5.7
ABS, mg/l	0.14	0.12
COD, mg/l	18	14

---

(1) Virgin 8 x 30 SGL Pittsburgh Carbon.

Effect of Flow Rate on Adsorption. PCC 2 and 4 were operated at surface loading rates of 4.1 and 6.5 gpm/ft^2, respectively. Figure 27 is a plot of the ratio of effluent ABS, C, to influent ABS, Co, as a function of residence time. The residence time was calculated using the empty volume of the carbon column. Figure 27 shows that flow rate has no significant effect on ABS adsorption by activated carbon. Figure 28 shows that flow rate did not significantly affect the adsorption of COD.

For the carbon system studied, flow rate, within the range of 4 to 6.5 gpm/sq ft did not significantly affect the capacity of the carbon. This conclusion is in agreement with the findings of Joyce, et al,(14,15) who found flow rates in the range of 2.5 to 10 gpm/sq ft did not affect the performance of the carbon at a given contact time.

Effect of Particle Size on Adsorption. Previous studies(12) have shown that the rate of adsorption varies as the inverse square of the diameter of the adsorbent particle. Since carbon particle size primarily affects the rate of adsorption and not the capacity of the carbon, the difference in performance of columns containing different size carbons decreases as the contact time increases.(13)

In order to determine the benefit of using a smaller carbon particle size, two pilot carbon columns containing different sizes of carbon were

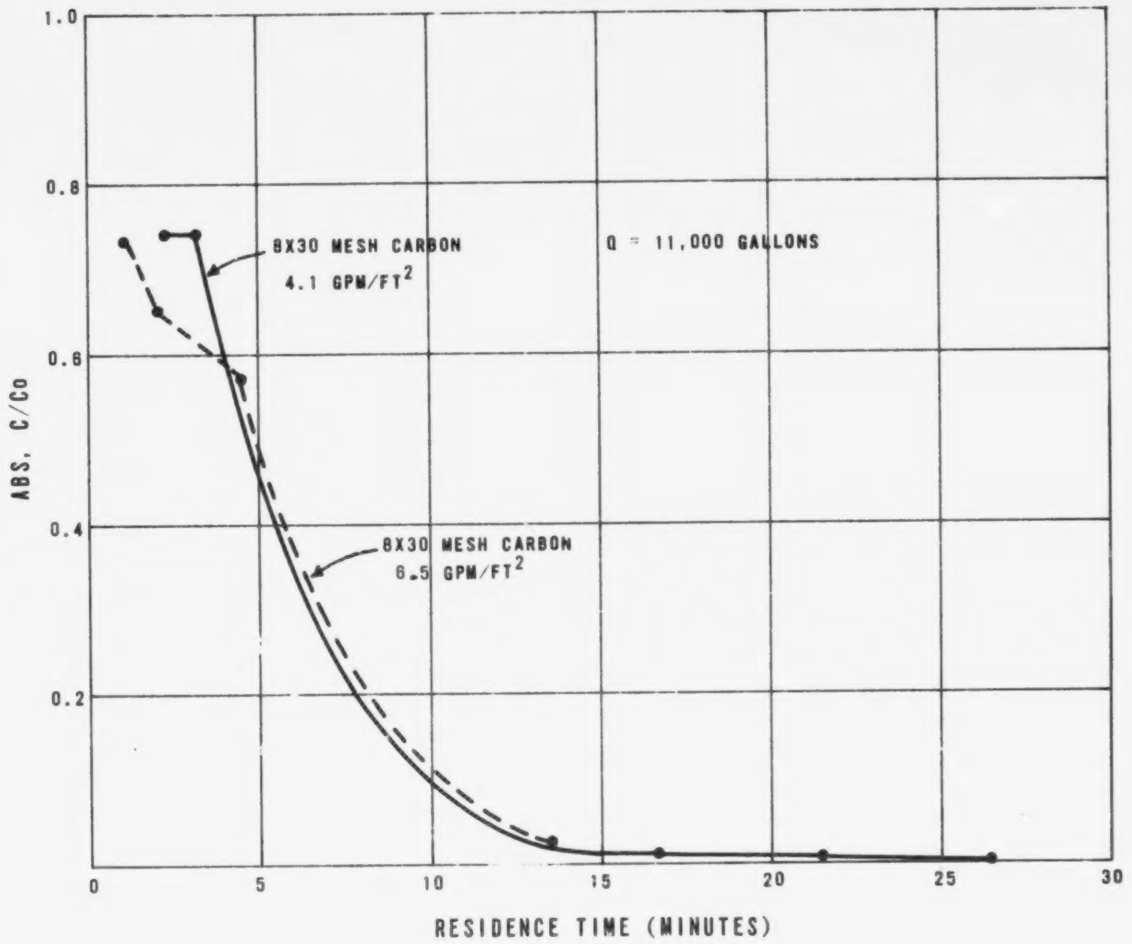


FIGURE 27

EFFECT OF FLOW RATE ON ADSORPTION

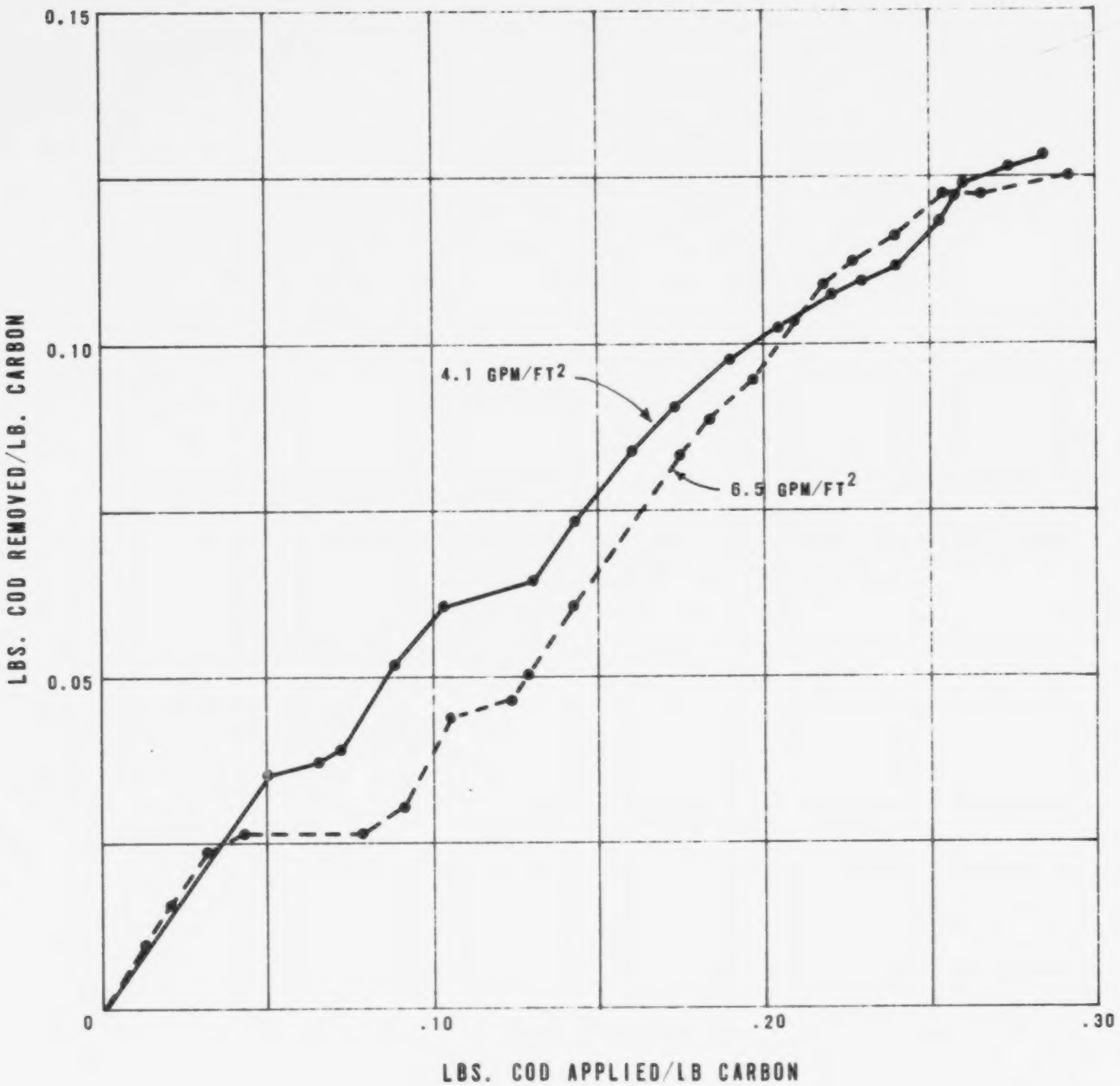


FIGURE 29

COMPARISON OF ADSORPTION OF 8X30 MESH ACTIVATED CARBON
AT A RESIDENCE TIME OF 7.8 MINUTES
AT TWO DIFFERENT FLOW RATES

operated in parallel. PCC 3 and 4 were, respectively, charged with 12 x 40-mesh and 8 x 30-mesh carbon and were operated at a surface loading rate of 6.5 gpm/sq ft.

The concentration curves of ABS, COD, and TOC for PCC 3 and 4 (Figures 15—18) show that for carbon depths of 6.75 feet and greater, the effluent concentrations of ABS, COD, and TOC are considerably less for the 12 x 40-mesh carbon than for the 8 x 30-mesh carbon. At a throughput volume of approximately 14,000 gallons, the ABS concentration at a carbon depth of 9.25 feet in the 12 x 40-mesh column was 0.05 mg/l. However, a carbon depth of 14.5 feet of the 8 x 30-mesh carbon was required to obtain the same ABS concentration at the same throughput volume.

A comparison of TOC and COD adsorption characteristics of the 12 x 40- and 8 x 30-mesh carbons is shown on Figures 29 and 30. Sufficient adsorption had not occurred above the 9.25-foot level to determine the quantitative difference, if any, in adsorptive capacity at greater carbon depths and longer residence times. It was planned to run PCC 3 (12 x 40-mesh carbon) to obtain data at these longer contact times, but the head loss across this column was so excessive that no water could be passed through the column, and the study had to be abandoned.

The study indicates that the 12 x 40-mesh carbon is more efficient in removing ABS, COD, and TOC than is the 8 x 30-mesh carbon for contact times less than 13.5 minutes. At a contact time of 10 minutes (9.25 feet — Figure 29), the 12 x 40-mesh carbon is about 35 percent more efficient than the 8 x 30-mesh carbon. Figure 31 shows a superior profile for the smaller carbon particle size. Future studies should include longer contact times to better evaluate the effect of carbon particle size on adsorption at longer contact times.

As previously discussed, the problem of excessive head loss across the pilot carbon columns could not be corrected. However, it was observed that the problems of excessive fines and physical fouling of the carbon were much more pronounced with the 12 x 40-mesh carbon than the 8 x 30-mesh carbon. Changes in pilot column design could reduce the head loss problem associated with both the 12 x 40-mesh and 8 x 30-mesh carbons in the pilot columns.

Efficiency of Regenerated Carbon. Following the first regeneration of carbon, a pilot column of virgin 8 x 30-mesh carbon (PCC 5) was operated in parallel with a column of regenerated 8 x 30-mesh carbon (PCC 6). These columns were operated at a surface loading rate of 4.6 gpm/sq ft. Figures 32—34 show a comparison of the virgin and regenerated ABS profiles for carbon depths of 4.25, 6.75, and 9.25 feet. These profiles show that the regenerated carbon is not as efficient in removing ABS as is the virgin carbon.

A mathematical integration of the ABS breakthrough curves for PCC 5 and 6 was performed for carbon depths of 1.75, 4.25, and 6.75 feet. Since

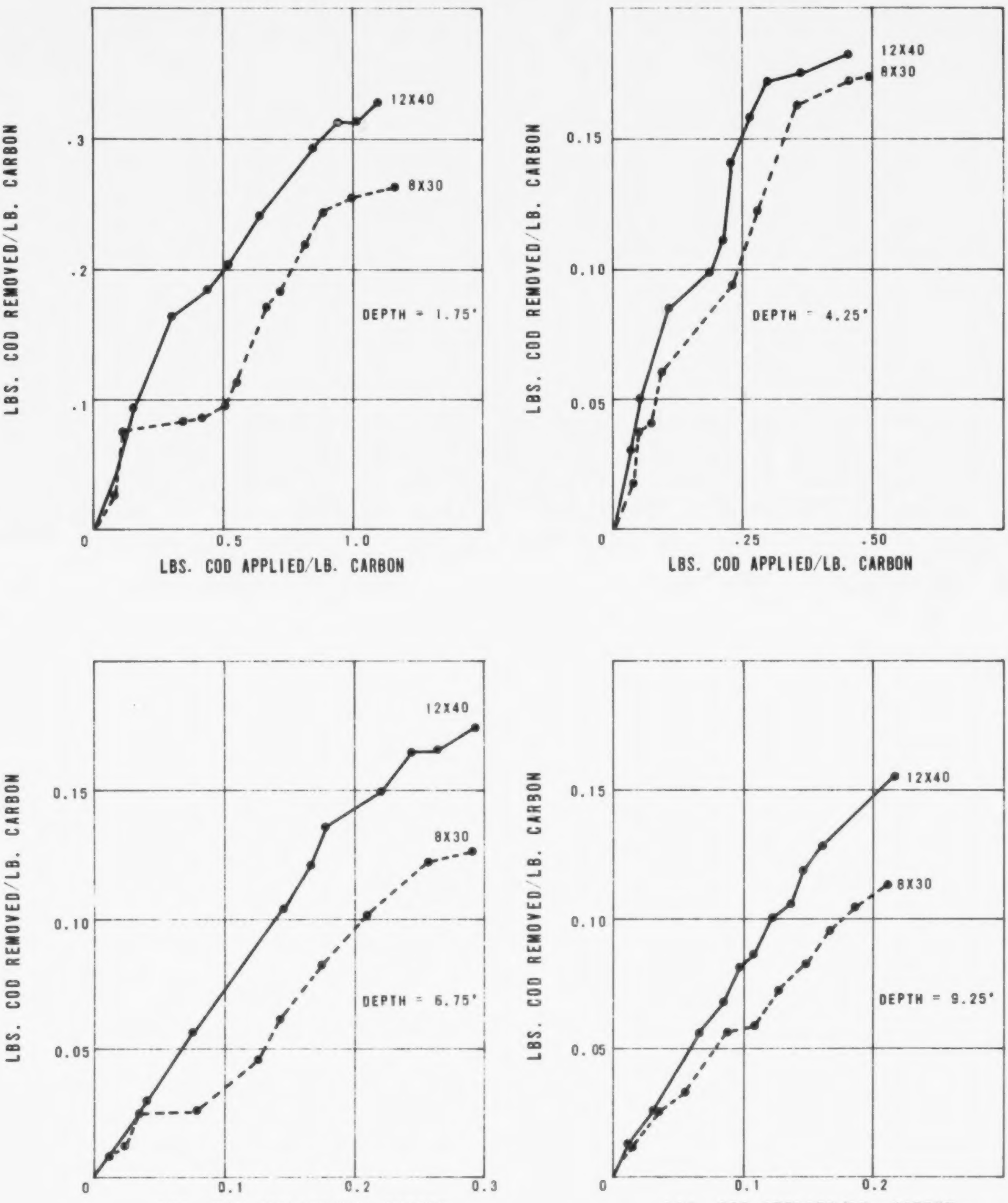


FIGURE 29

COMPARISON OF COD ADSORPTION BY 8X30 AND 12X40 MESH ACTIVATED CARBON

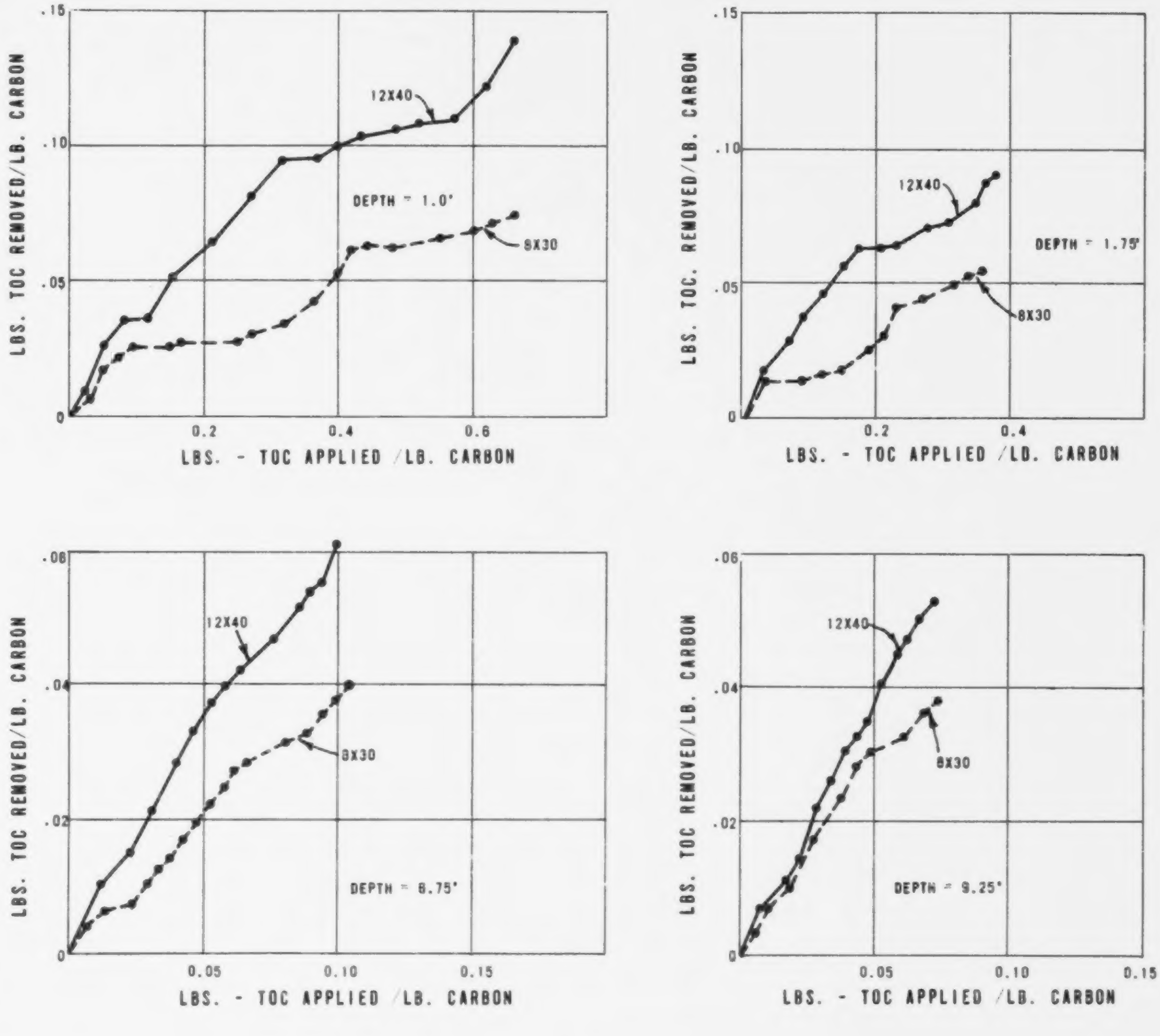


FIGURE 30
COMPARISON OF TOC ADSORPTION
BY 8X30 AND 12X40 MESH
ACTIVATED CARBON

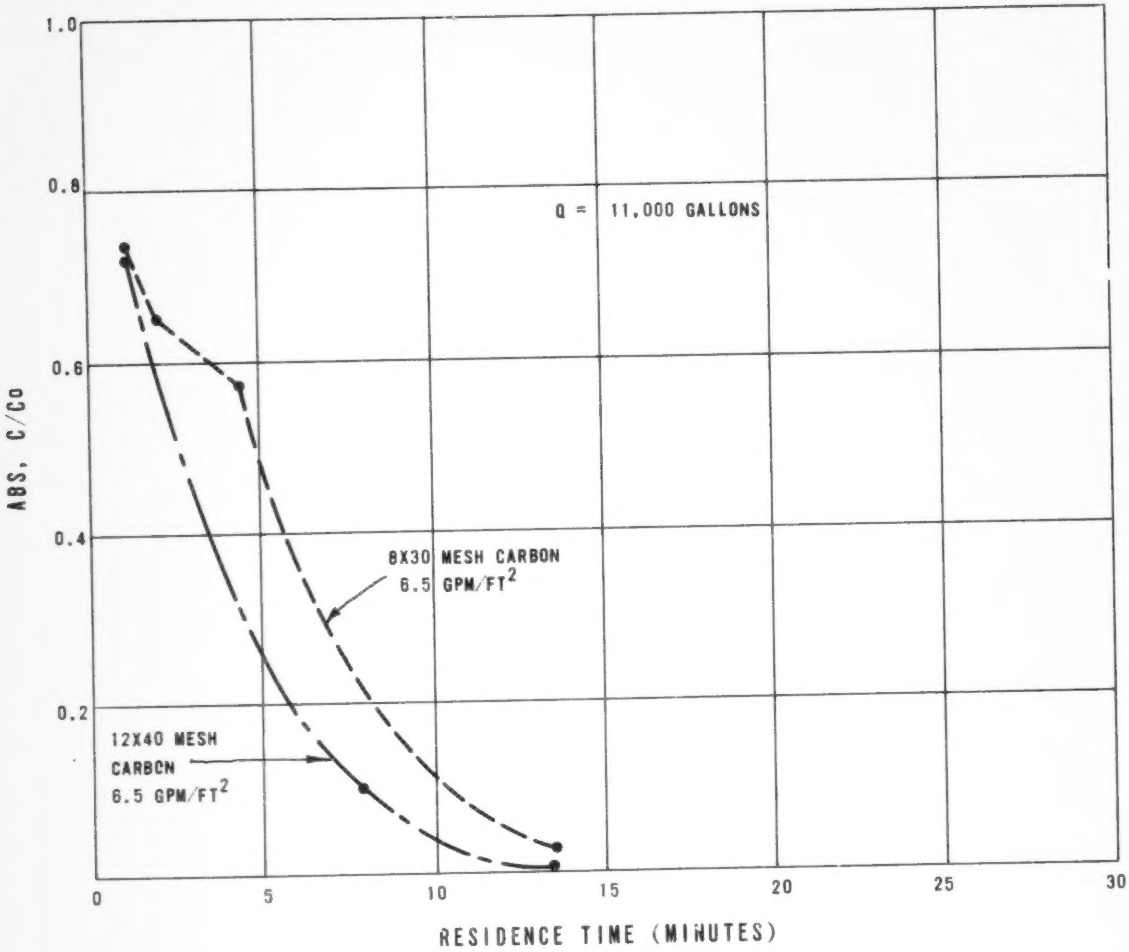


FIGURE 31

EFFECT OF CARBON PARTICLE SIZE ON ADSORPTION

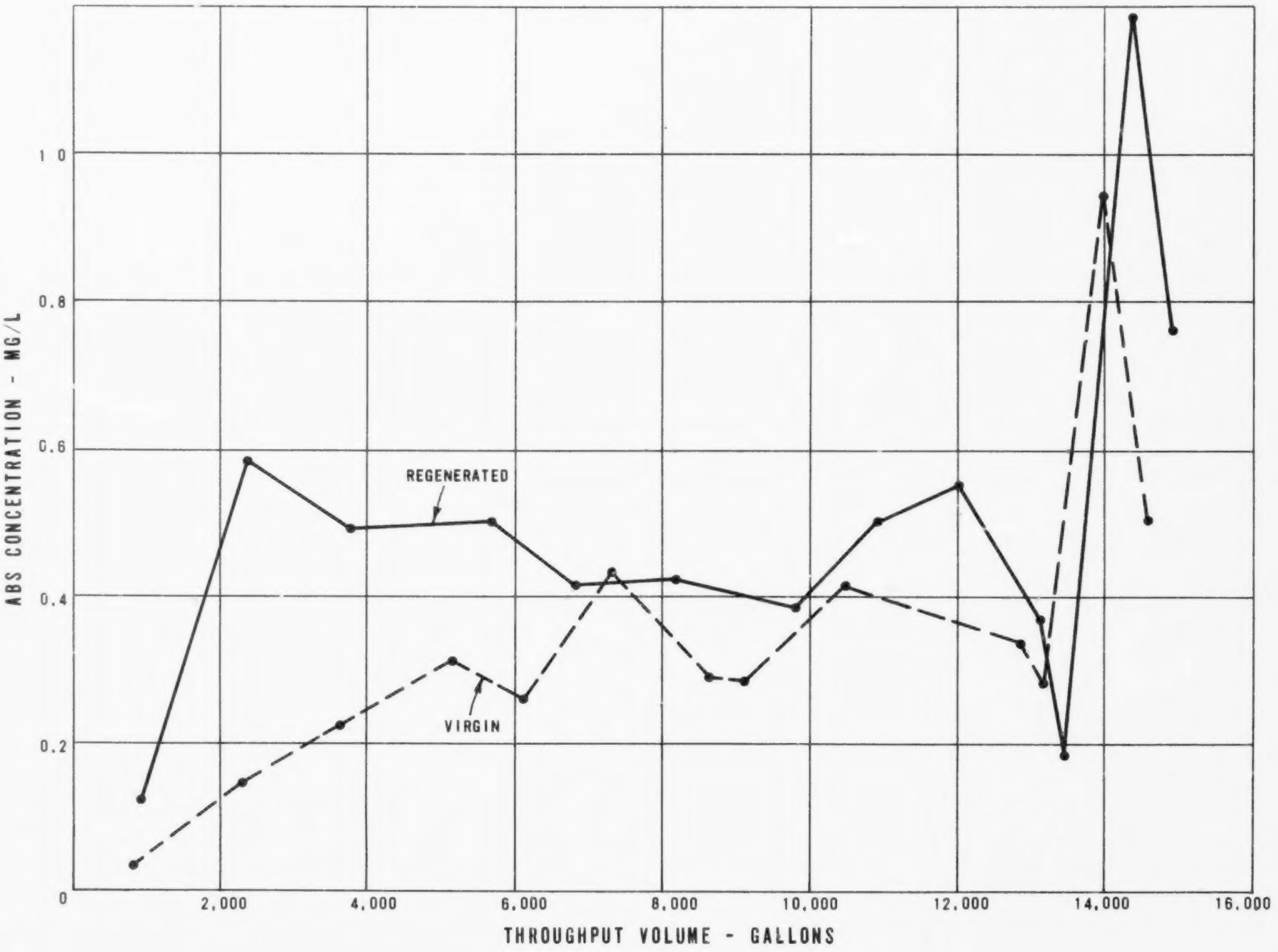


FIGURE 32
COMPARISON OF VIRGIN AND 1st CYCLE REGENERATED CARBON
ABS PROFILE FOR 4.25 FT. CARBON DEPTH

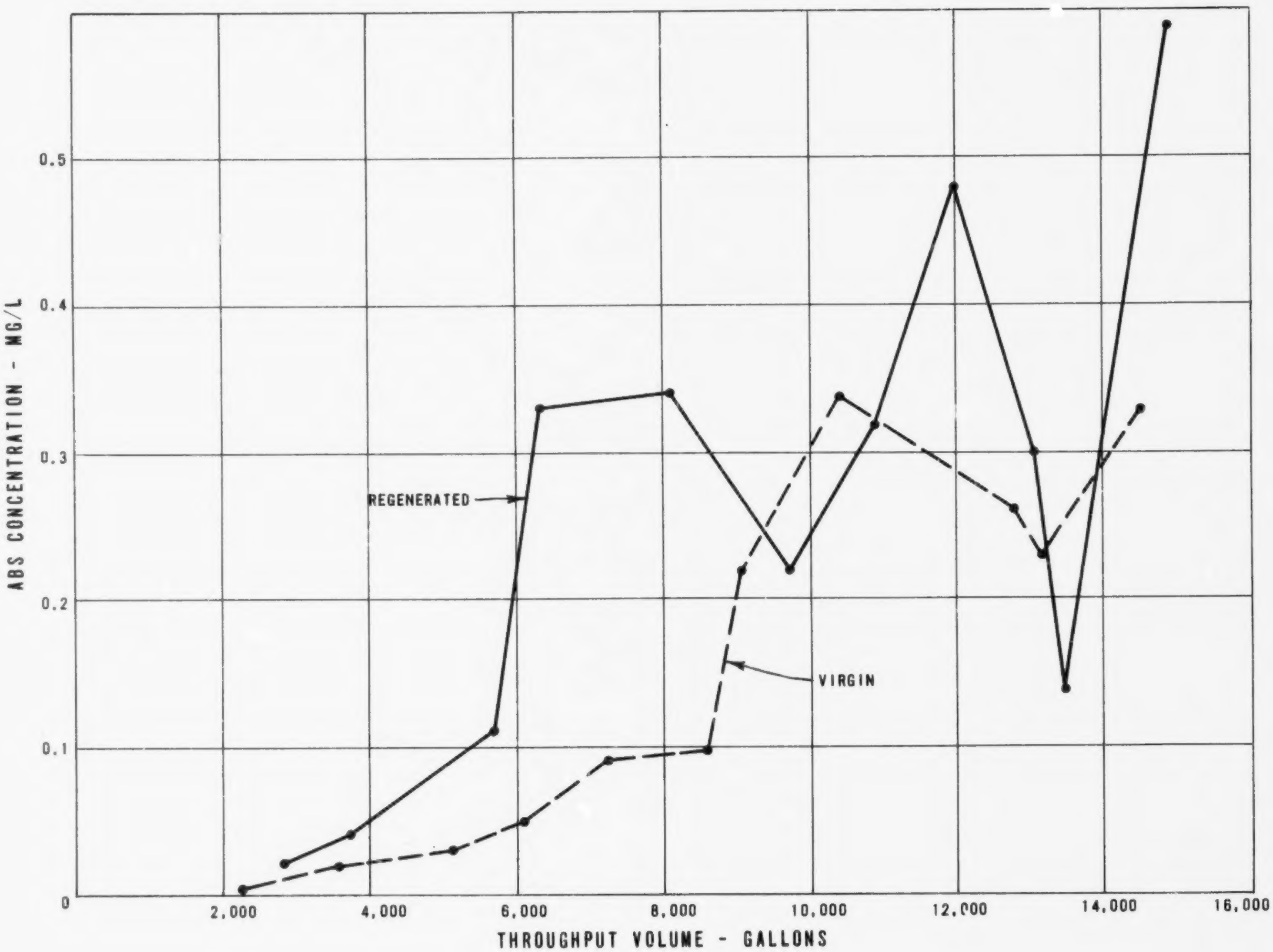


FIGURE 33

COMPARISON OF VIRGIN AND 1ST CYCLE REGENERATED CARBON
ABS PROFILE FOR 6.75 FT. CARBON DEPTH

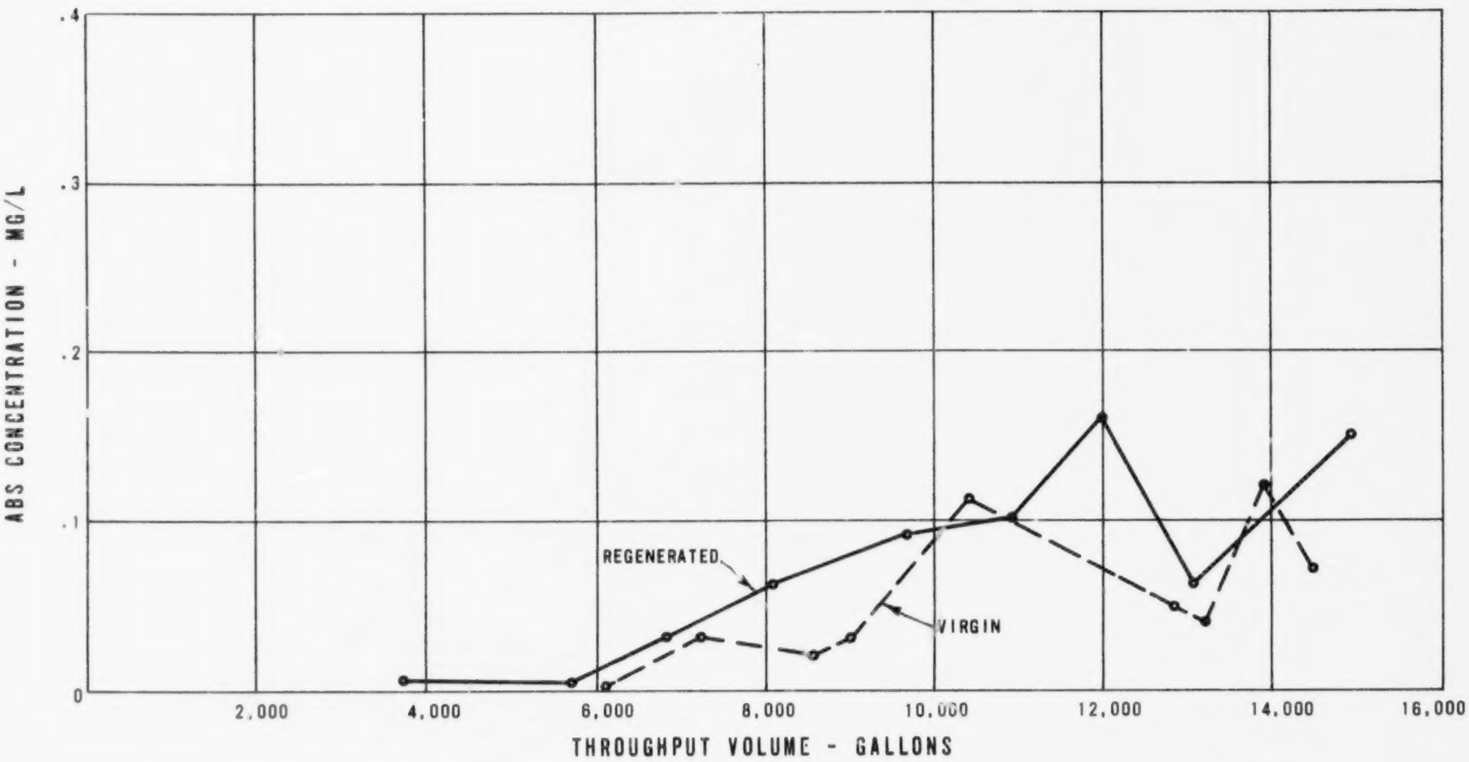


FIGURE 34
COMPARISON OF VIRGIN AND 1ST CYCLE REGENERATED CARBON
ABS PROFILE FOR 9.25 FT. CARBON DEPTH

the columns were charged with identical quantities of carbon and operated at identical surface loading rates, the results were expressed as pounds of ABS removed vs. pounds of ABS applied (Figures 35—37). The regenerated carbon can be seen to be less efficient in removing ABS.

Figure 38 indicates that the regenerated carbon is at least 10 percent less efficient in removing ABS than is the virgin carbon for contact times less than 11 minutes. The main columns have not shown an increased carbon requirement, however, due to the lower ABS concentration.

Experimental Results of Run 3

The data from Run 3 of the upflow columns are presented in Figures 39—54. The influent and effluent concentrations of COD (chemical oxygen demand) are shown on Figure 39. The apparent decrease of influent concentration throughout the run is probably due to the combination of a more dilute winter sewage and more efficient chemical clarification. The adsorption efficiency in the columns decreases as the influent concentration decreases.

Figures 40—43 show the COD concentrations at four depths (contact times) for each of the pilot carbon columns. It should be noted that although PCC 15 was still removing COD, the head loss in the column became excessive, and the column was shut down.

A comparison of the four carbons tested at three bed depths is made in Figures 44—46. Pounds of COD removed per pound of carbon vs. the amount of COD applied is used as the basis for comparison. Similar trends are noticed in each of the figures, but more difference between the carbons is apparent with increasing bed depth. In the lower bed depths (Figures 44 and 45), the Brand C carbon appears to have reached a saturation value of about 0.12 pounds COD per pound. The other carbons did not appear to have reached a saturation value at the end of the run. Figure 46 illustrates the greatest difference between the carbons used. After 0.25 pounds COD per pound of carbon were applied, the Pittsburgh SGL carbon had removed twice as much as the Brand B carbon and a third again as much as the Brand A carbon.

Figures 47, 48, and 49 show the ratio of effluent COD concentration (C) to influent COD concentration (Co) at three bed depths. It can be seen from Figure 47 that there is little difference between the carbons in ability to removed COD at a short contact time. The difference becomes more pronounced, however, at longer contact times (Figures 48 and 49). Figure 50 also illustrates this fact. It is apparent that the Pittsburgh Type SGL achieved the best removal of all the carbons tested in this run, using filtered secondary effluent at South Lake Tahoe. After about 7 pounds of COD had been applied to each column, the removals had averaged 76 percent for SGL, 56 percent for Brand A, and 37 percent for Brand B.

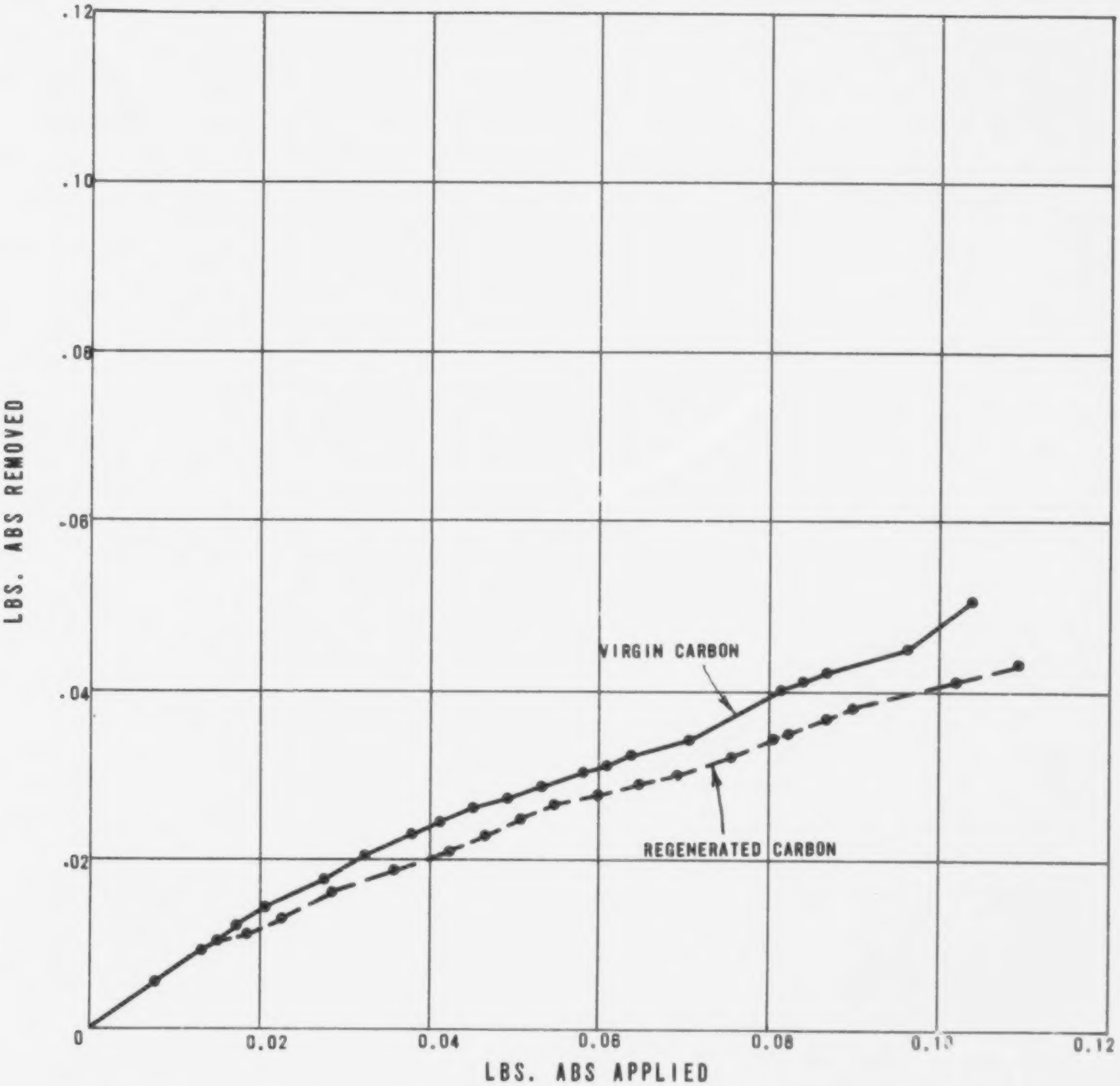


FIGURE 35
ABS ADSORPTION FOR
VIRGIN AND REGENERATED CARBON
1.75' CARBON DEPTH, 2.8 MINUTE CONTACT TIME

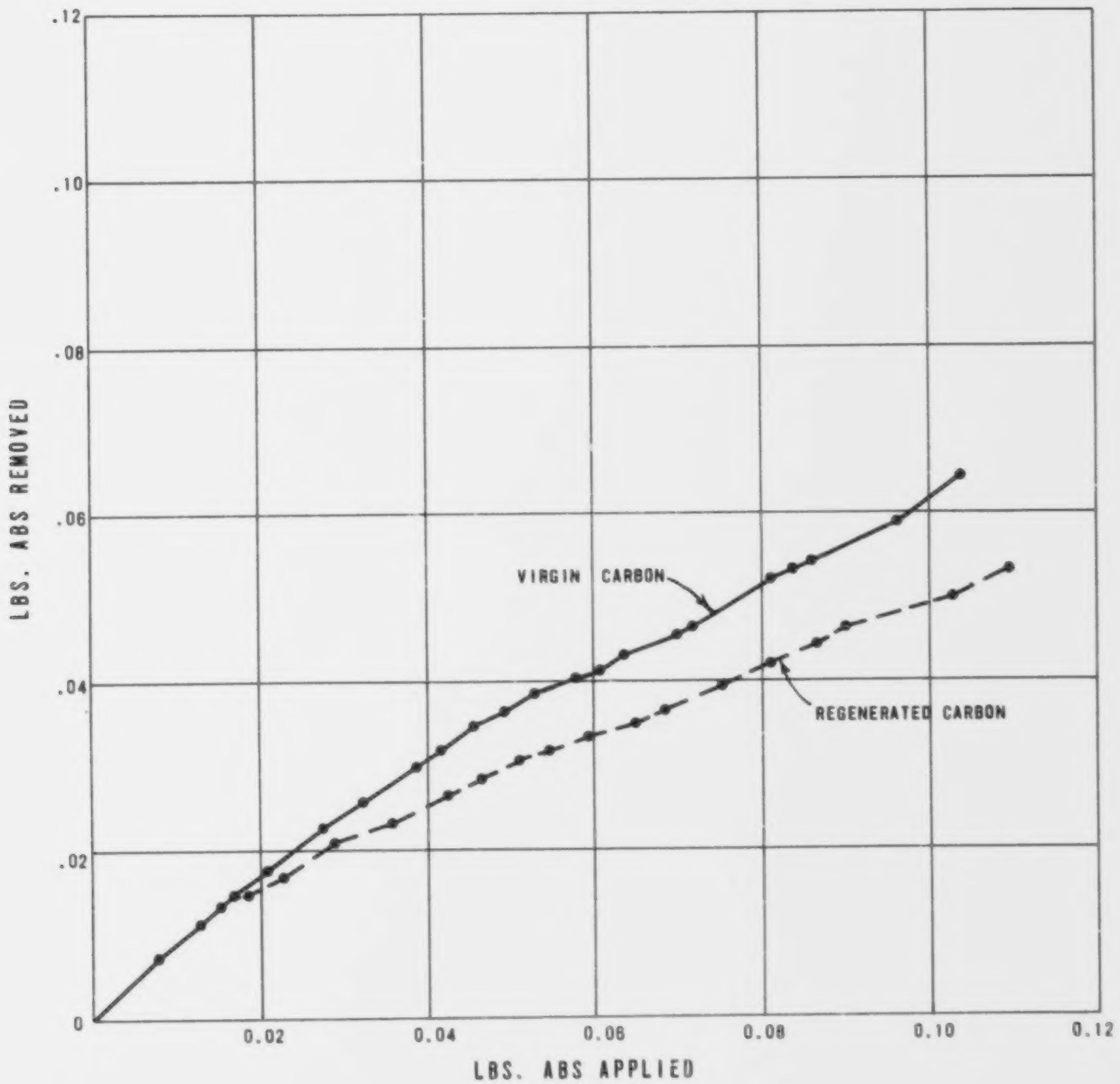


FIGURE 36
ABS ADSORPTION FOR
VIRGIN AND REGENERATED CARBON
4.25' CARBON DEPTH, 6.9 MINUTE CONTACT TIME

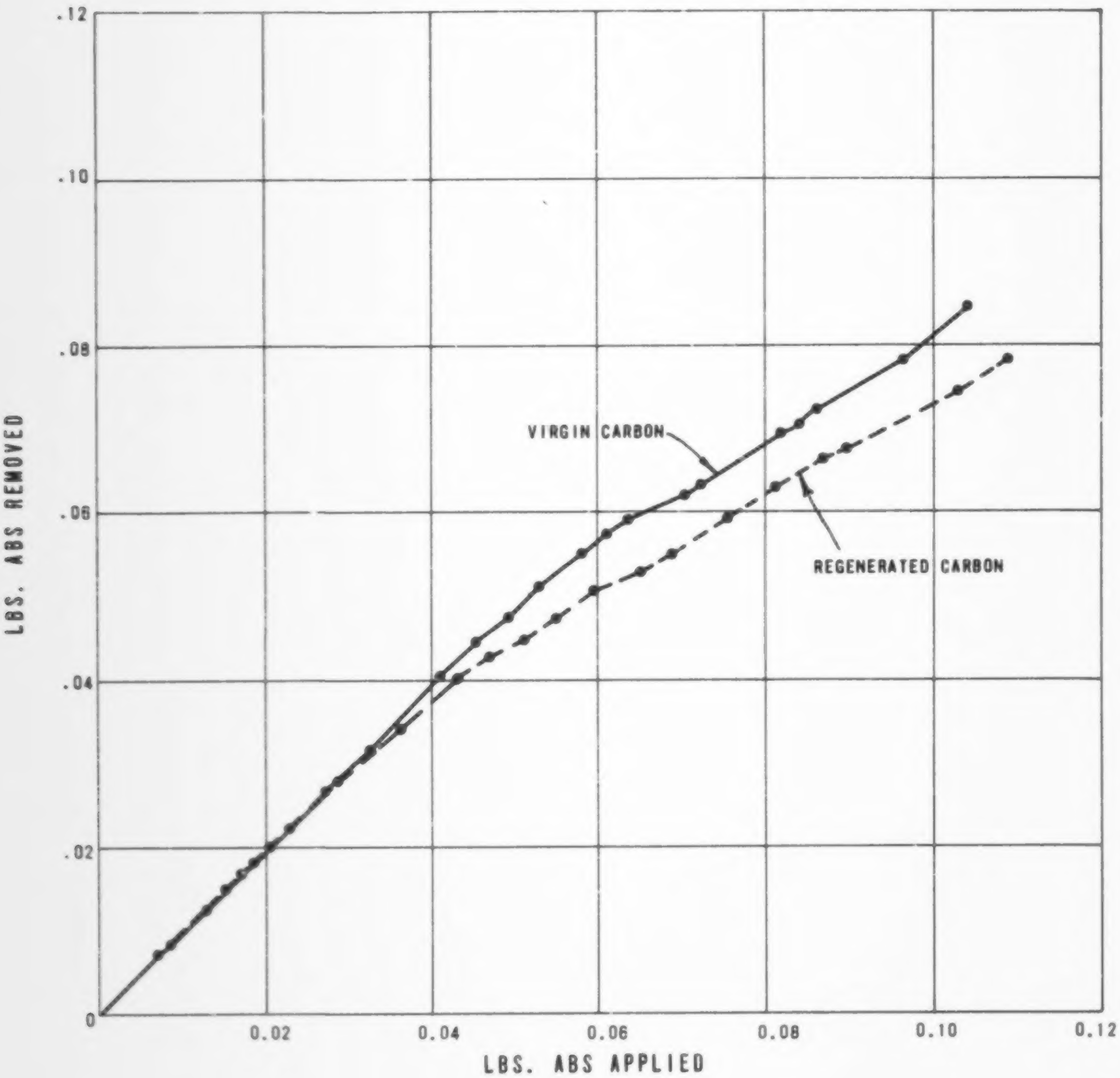


FIGURE 37
ABS ADSORPTION FOR
VIRGIN AND REGENERATED CARBON
6.75' CARBON DEPTH, 11 MINUTE CONTACT TIME

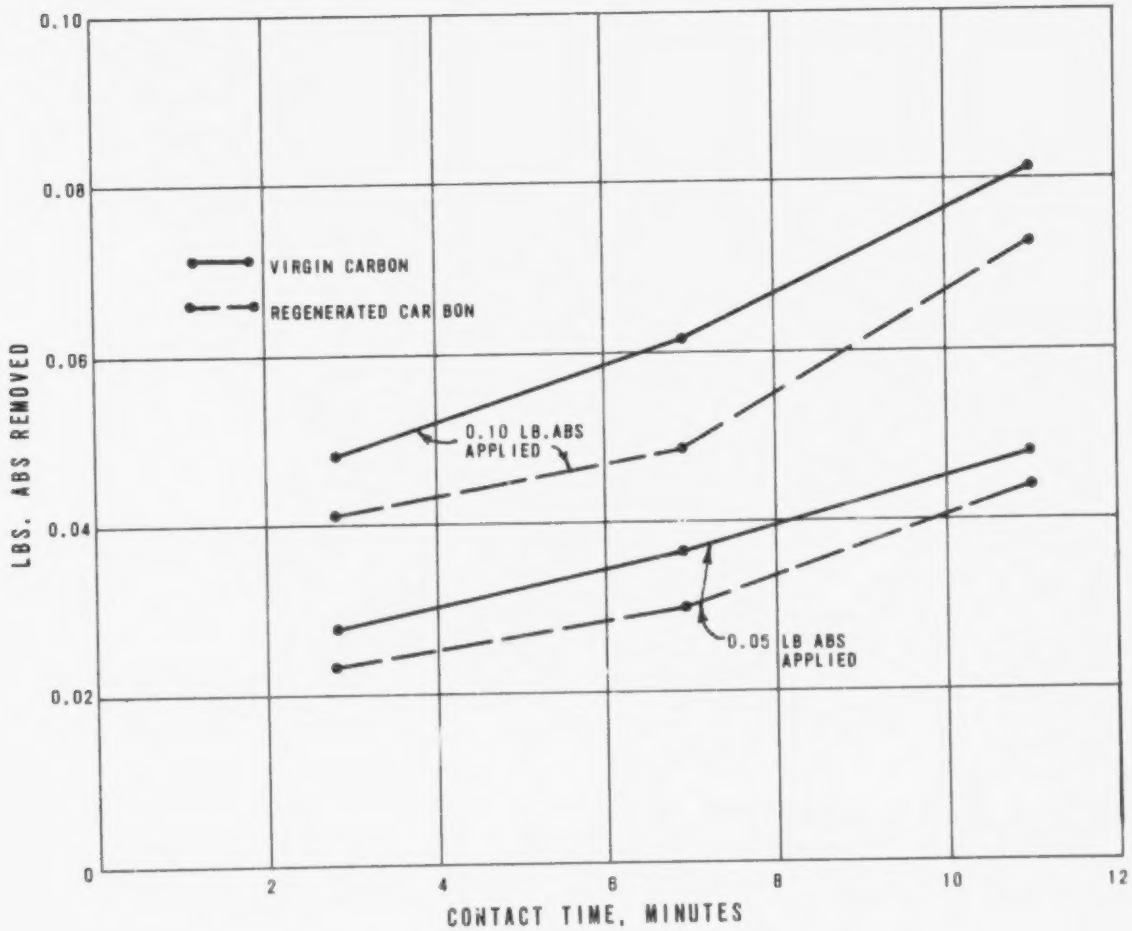
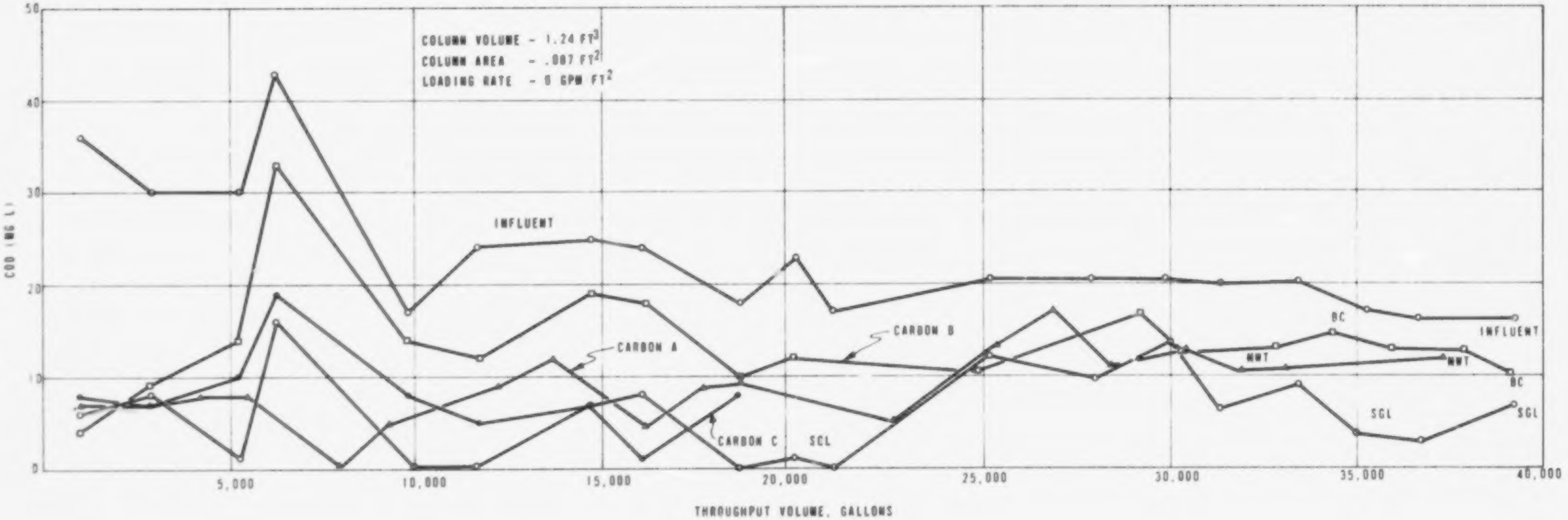


FIGURE 38

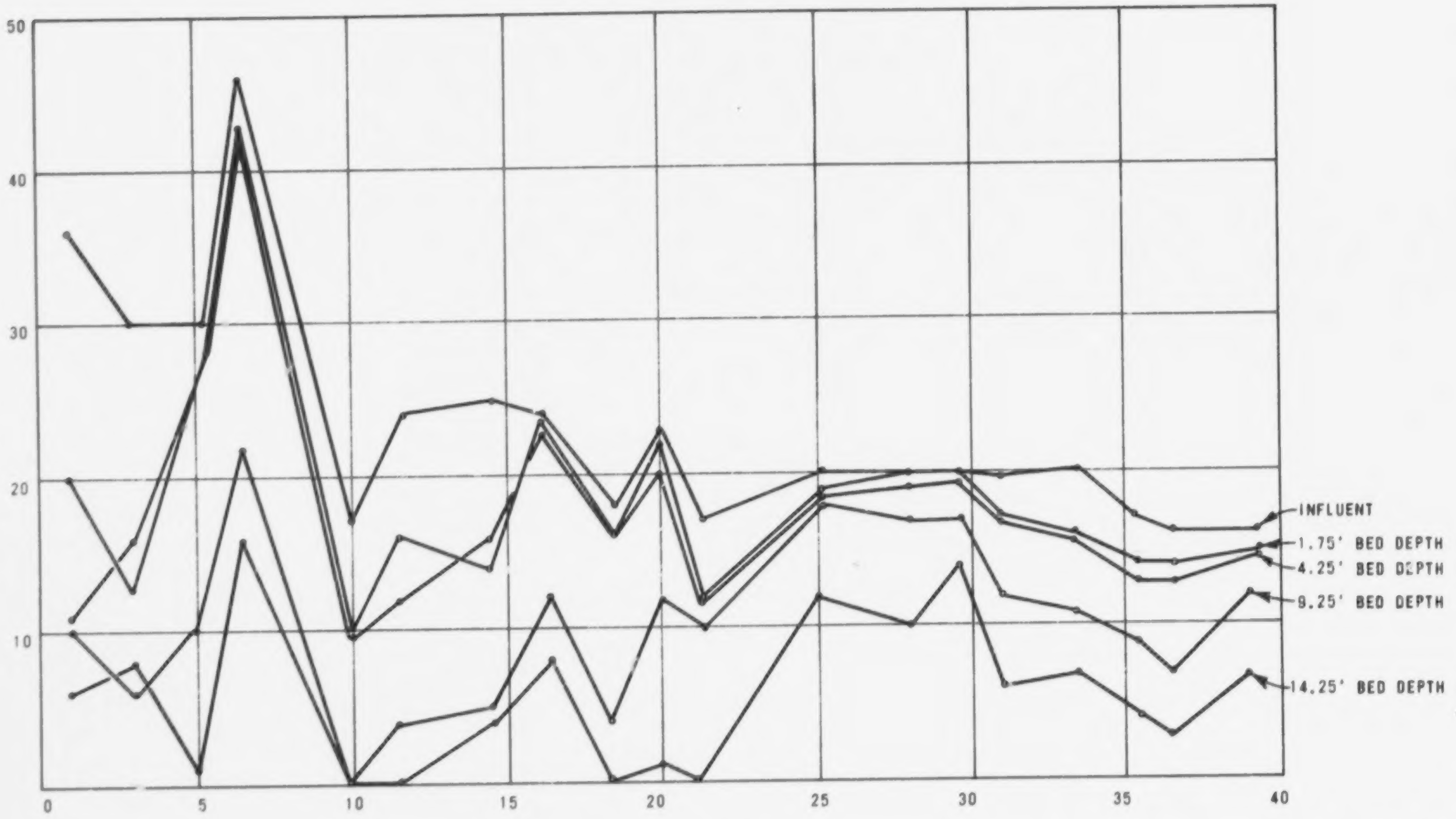
COMPARISON OF VIRGIN AND 1ST CYCLE REGENERATED CARBON FOR ADSORPTION OF ABS AS A FUNCTION OF CONTACT TIME



THROUGHPUT VOLUME, GALLONS

FIGURE 39

COMPARISON OF C O D PROFILES FOR PILOT CARBON COLUMNS AT 14.25 FEET BED DEPTH
AND 17.5 MINUTE RESIDENCE TIME



GALLONS PASSED X 1,000

FIGURE 40

C.O.D. CONCENTRATION HISTORY AT FOUR BED DEPTHS FOR
PCC 12 - UPFLOW COLUMN CHARGED WITH PITTSBURGH
TYPE SGL ACTIVATED CARBON

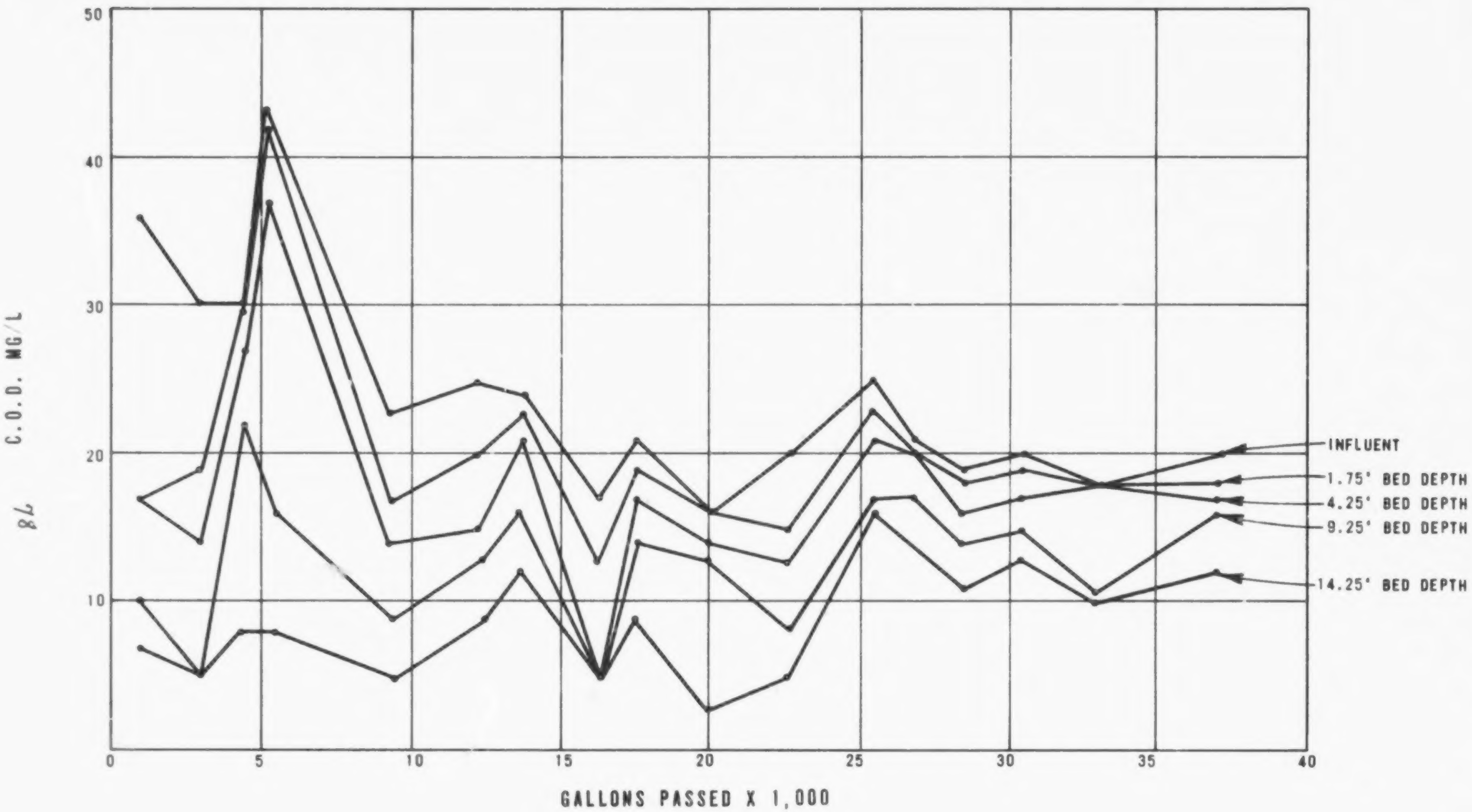


FIGURE 41

C.O.D. CONCENTRATION HISTORY AT VARIOUS BED DEPTHS FOR PCC 13
UPFLOW COLUMN CHARGED WITH BRAND A CARBON

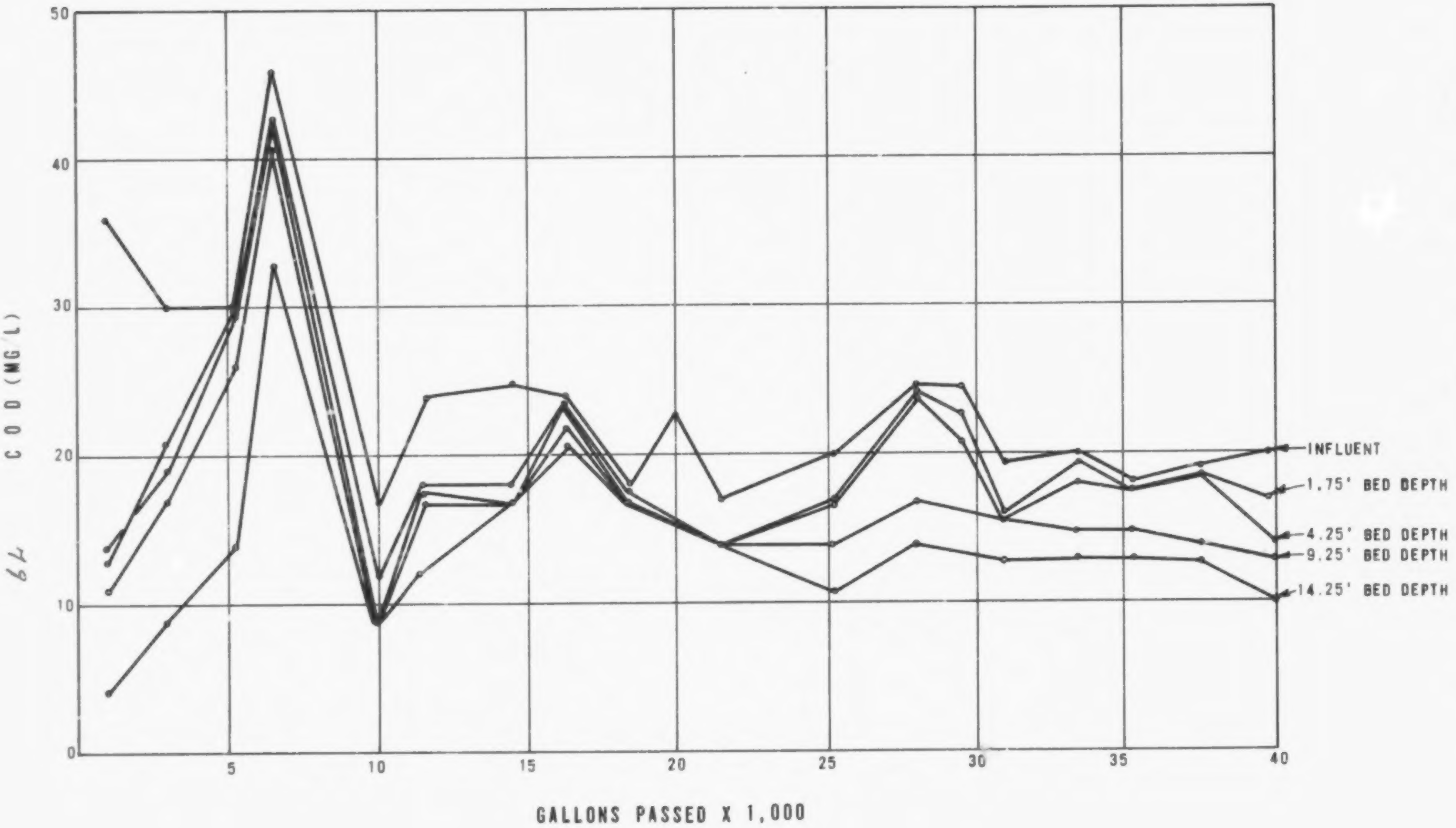


FIGURE 42

C O D CONCENTRATION HISTORY AT VARIOUS BED DEPTHS FOR PCC 14
UPFLOW COLUMN CHARGED WITH BRAND B ACTIVATED CHARCOAL

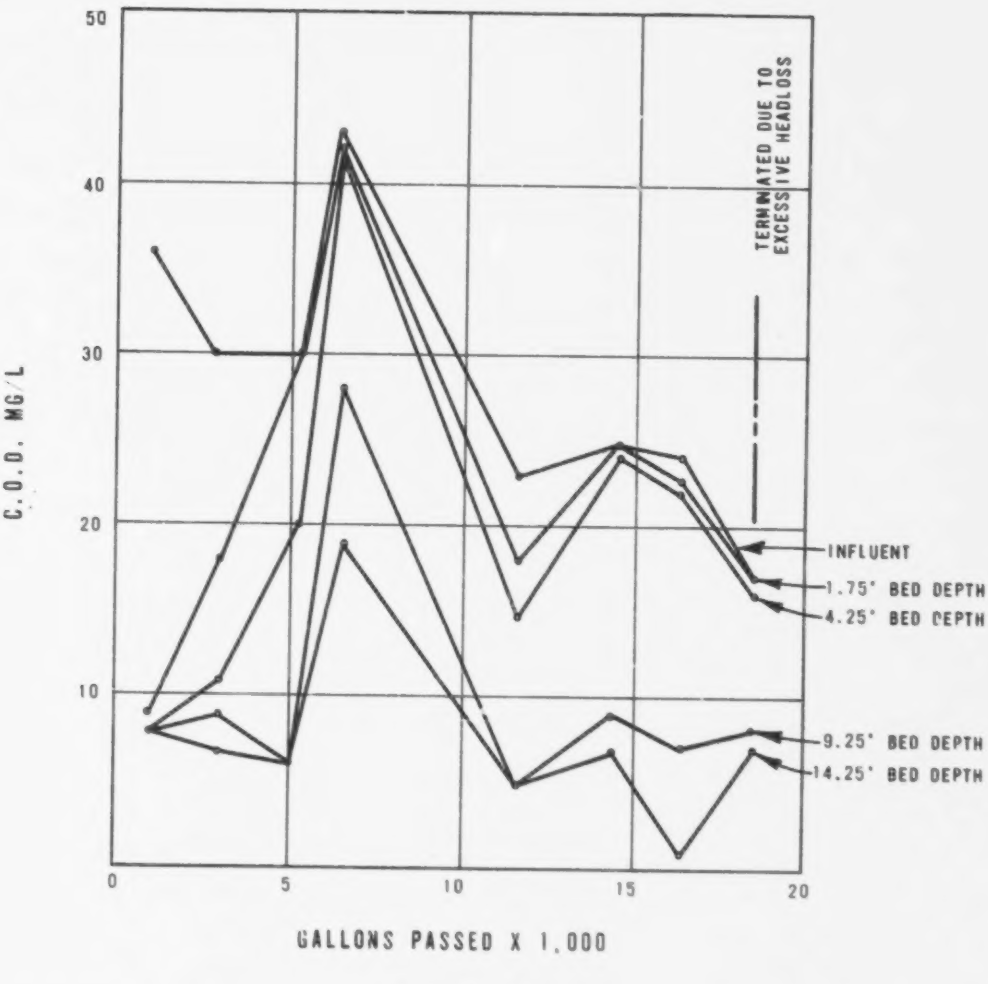
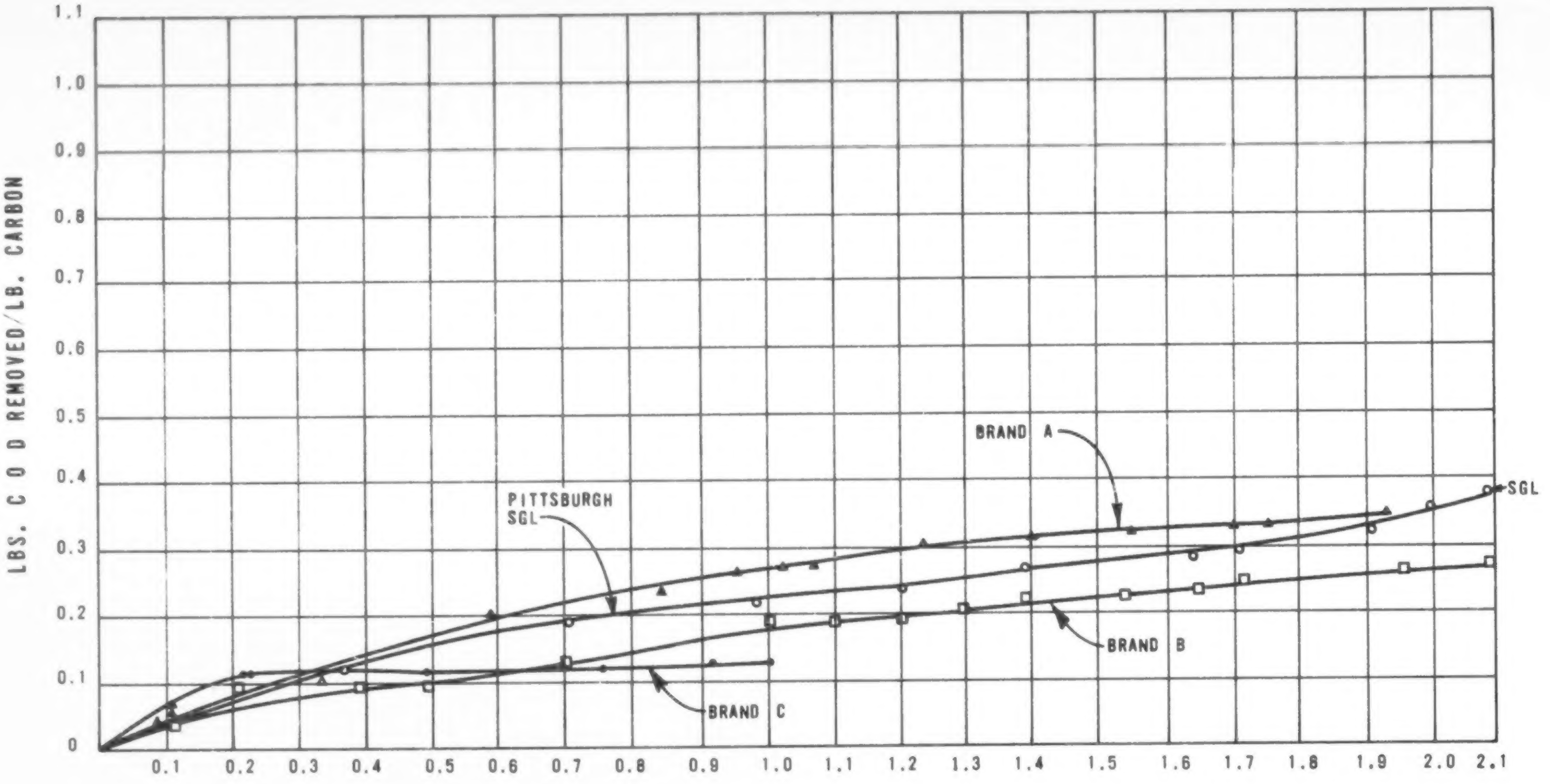


FIGURE 43
C.O.D. CONCENTRATION HISTORY AT VARIOUS BED DEPTHS FOR PCC 15
UPFLOW COLUMN CHARGED WITH TYPE C CARBON



LBS. C O D APPLIED/LB. CARBON

FIGURE 44

COMPARISON OF ADSORPTIVE CAPACITIES OF TEST CARBONS
1.75 FEET BED DEPTH - 2.2 MINUTE CONTACT TIME

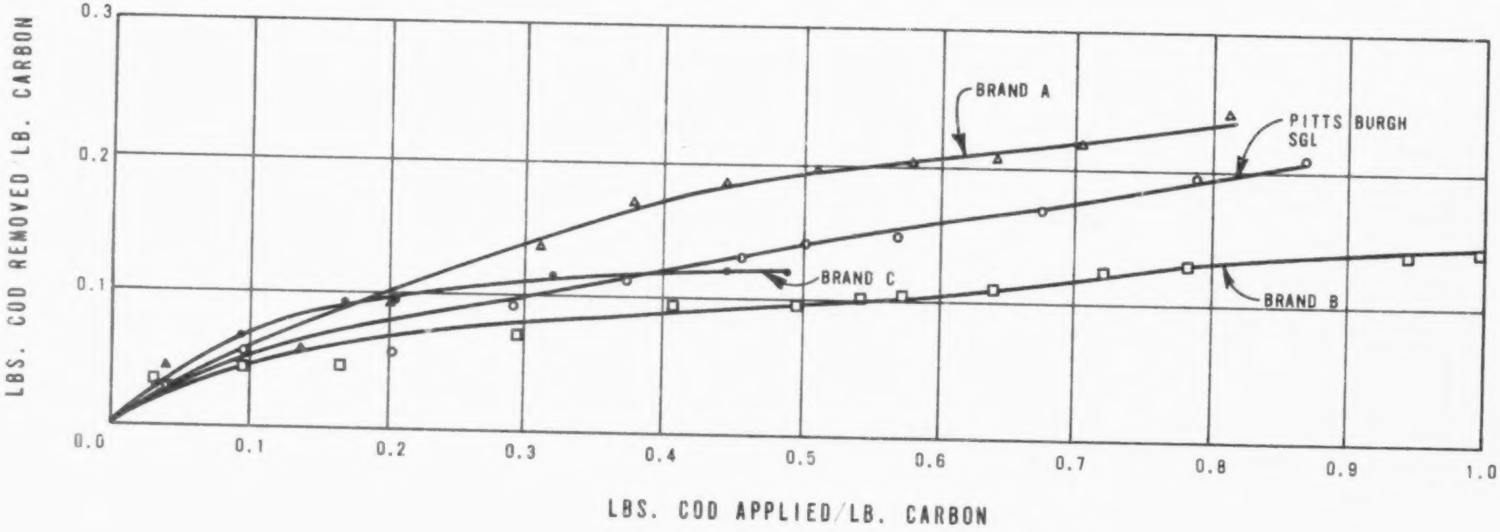


FIGURE 45

COMPARISON OF ADSORPTIVE CAPACITIES OF TEST CARBONS
4.25 FEET CARBON BED DEPTH - 4.6 MINUTE CONTACT TIME

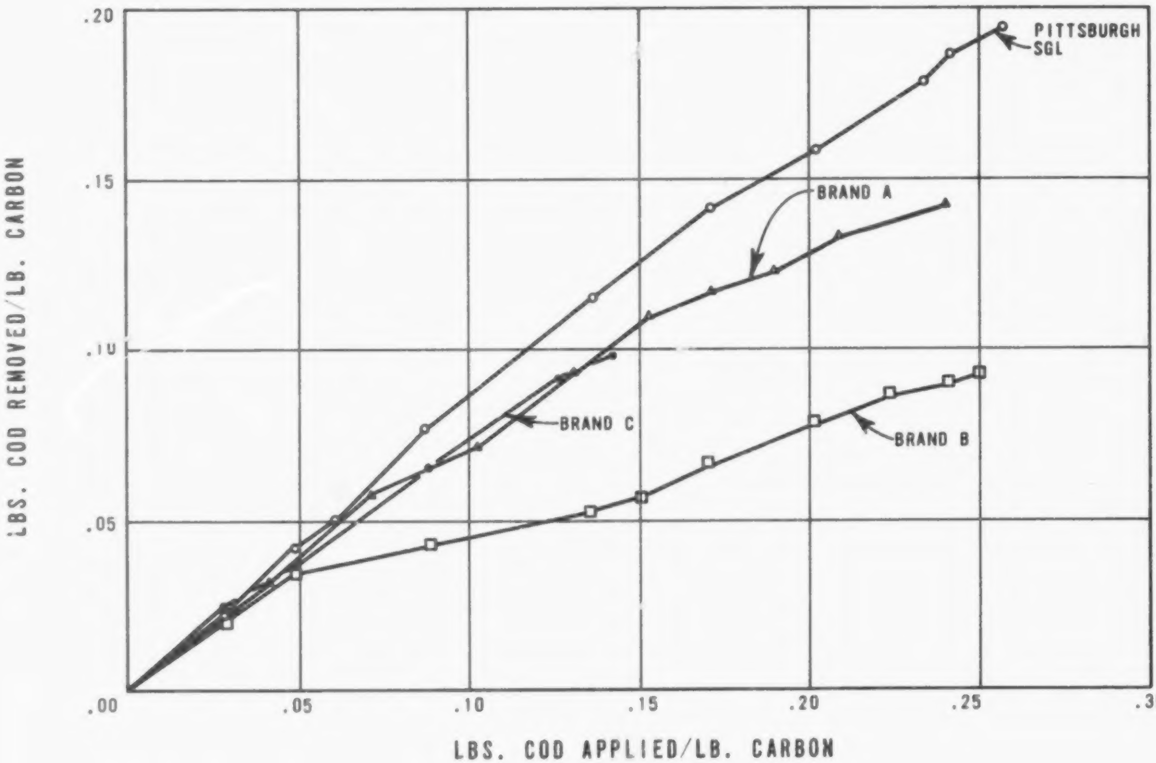


FIGURE 46

COMPARISON OF ADSORPTIVE CAPACITIES OF TEST CARBONS
14.25 FEET BED DEPTH - 17.5 MINUTE RESIDENCE TIME

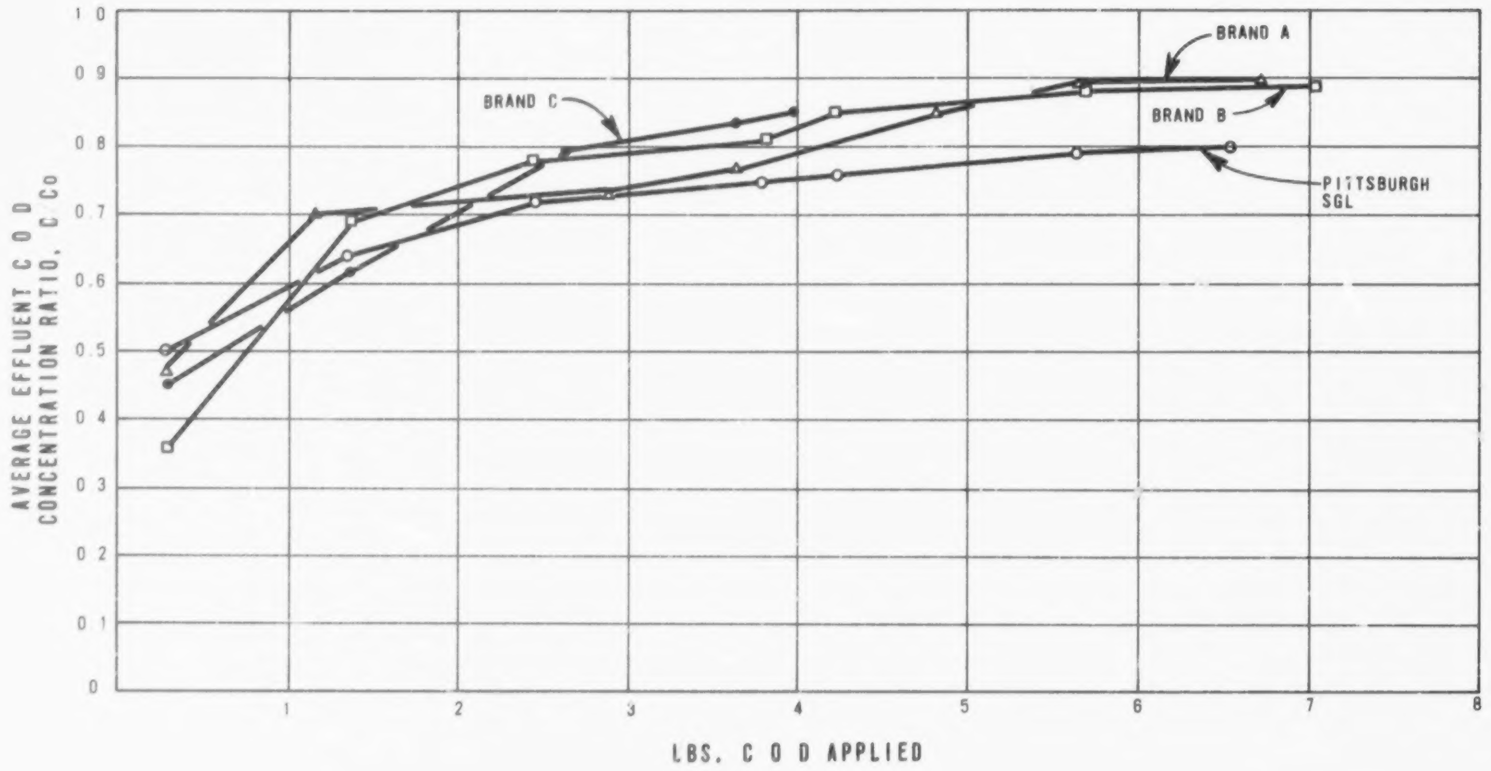


FIGURE 47

A COMPARISON OF THE C O D REMOVAL ABILITY OF THE TEST CARBONS
SAMPLE DEPTH - 1.75 FEET

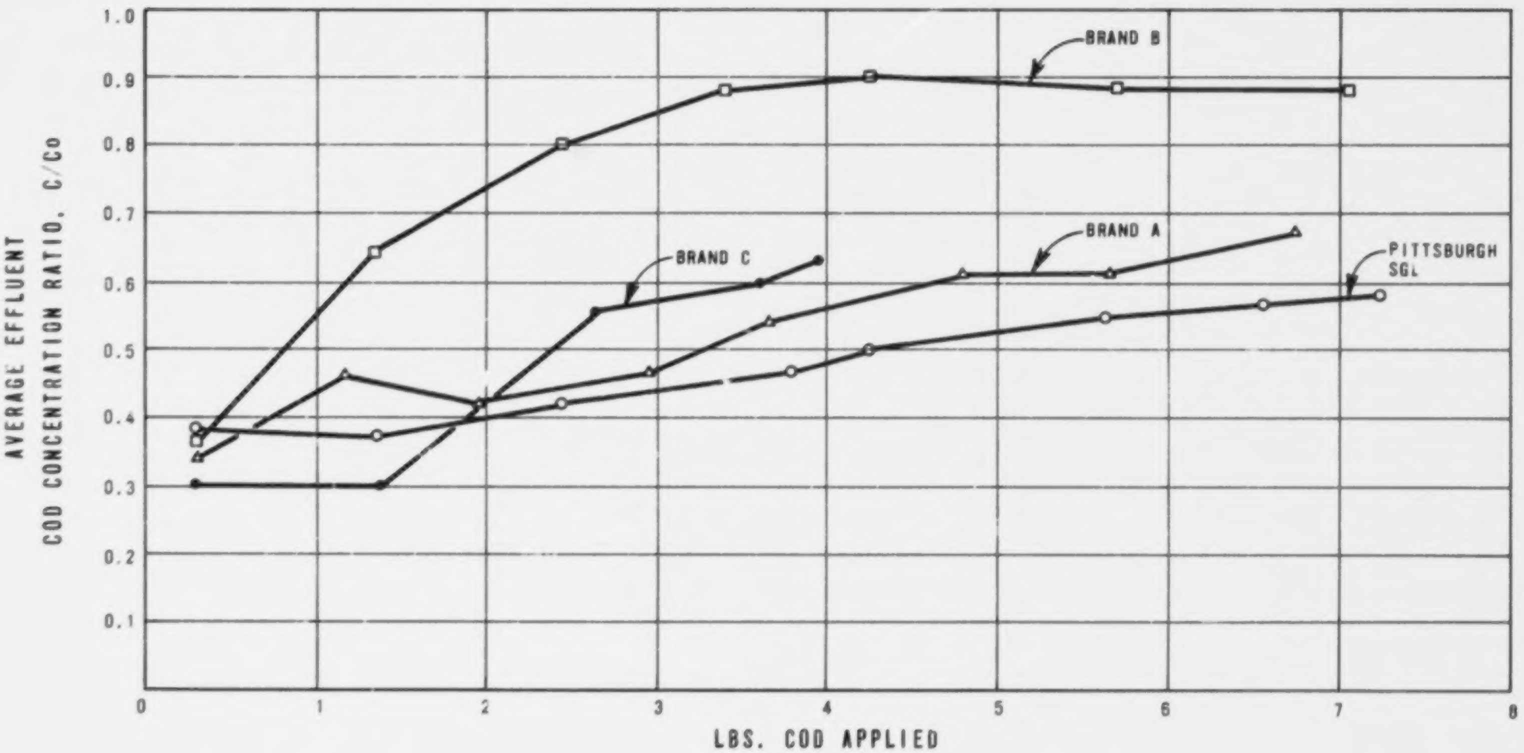


FIGURE 4B

A COMPARISON OF THE COD REMOVAL ABILITY OF THE TEST CARBONS
SAMPLE DEPTH - 6.75 FEET

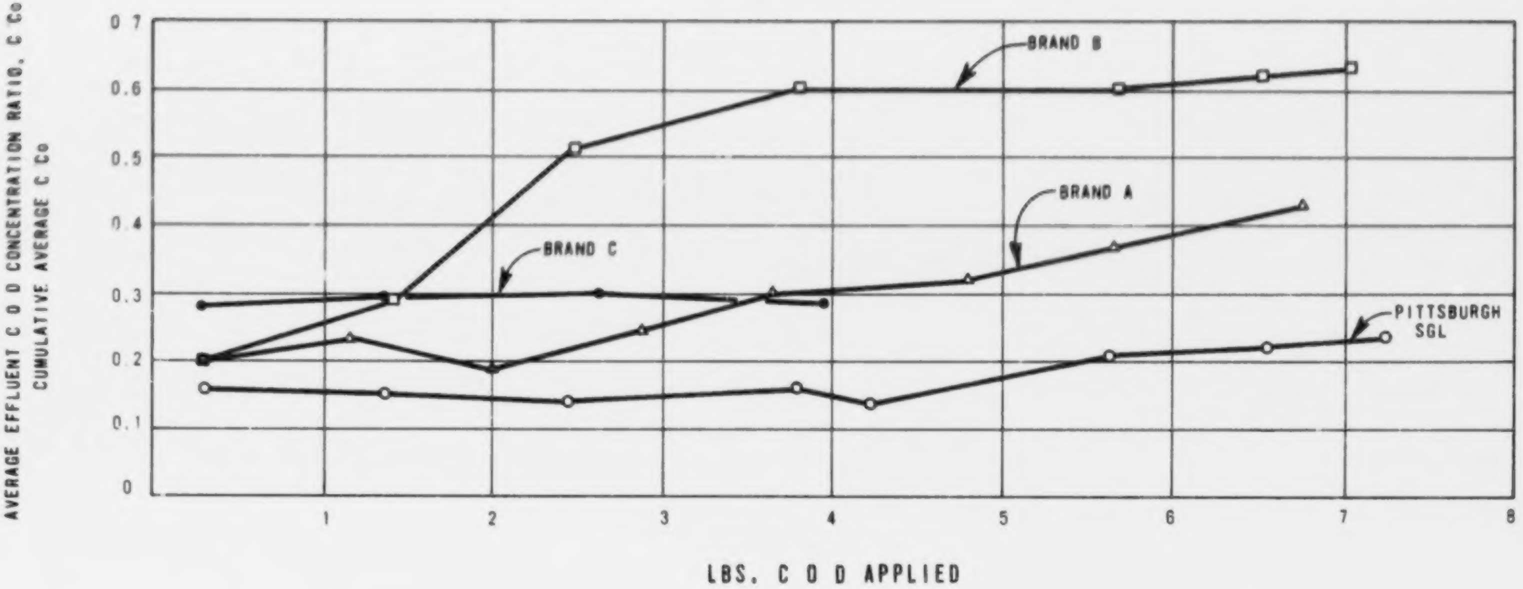


FIGURE 49

A COMPARISON OF THE C O D REMOVAL ABILITY OF THE TEST CARBONS

SAMPLE DEPTH – 14.25 FEET

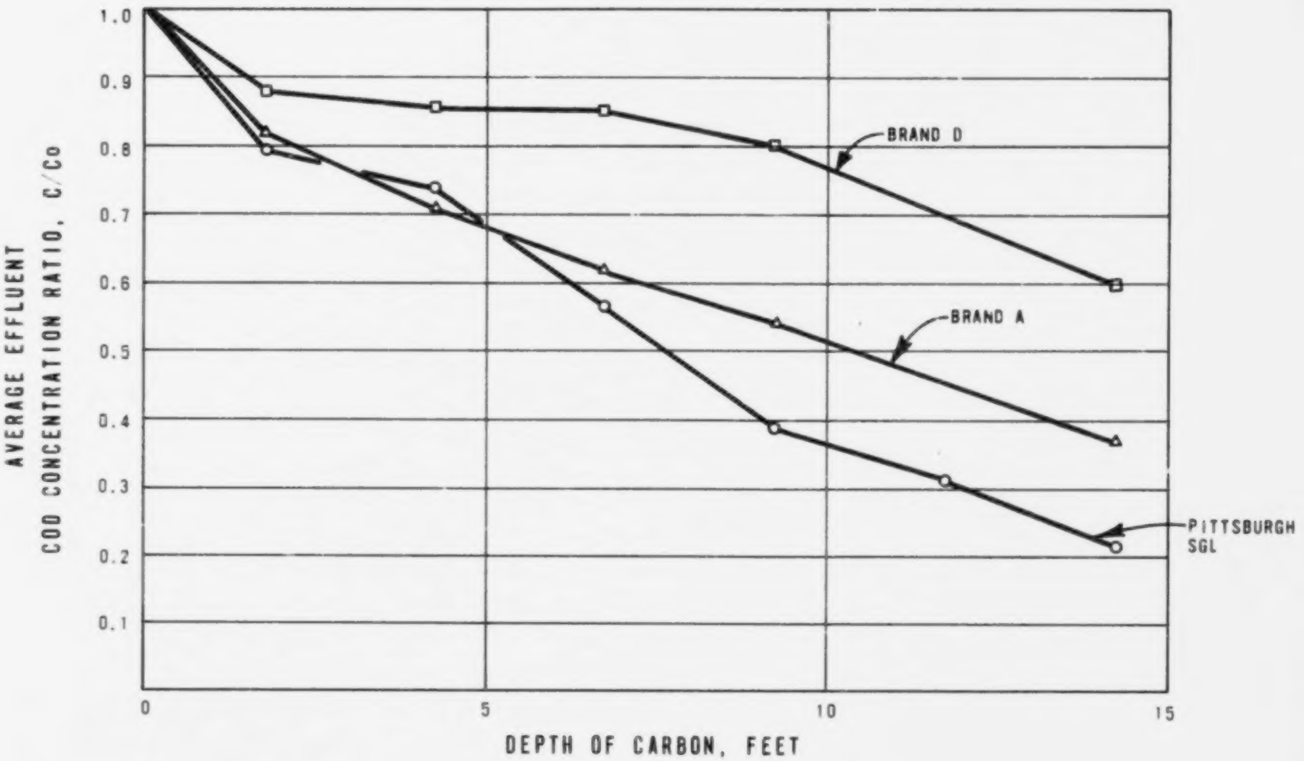


FIGURE 50

EFFECT OF CONTACT TIME ON COD REMOVAL
(APPROXIMATELY 0.2 LBS. COD APPLIED PER LB. CARBON)

The effect of contact time is further shown in Figure 51. The SGL Pittsburgh carbon, at a contact depth of only 9.25 feet, produced an effluent of the same quality as did the Brand A carbon after a 14.25-foot contact depth. At the end of the run, the SGL carbon removed 56 percent of the influent COD at the 9.25-foot bed depth. During this period, 14.25 feet of the Brand A carbon averaged 57 percent COD removal.

The conclusions drawn from this pilot column comparison of the four carbons include:

1. The sample of Brand C, 12 x 40-mesh activated carbon was shown to have the lowest saturation value (0.12 lbs removed/lb carbon) of the carbons tested.

2. The Brand C carbon also demonstrated a tendency for excessive head loss development. This problem may be related to the size and softness of the carbon.

3. The Brand B carbon sample tested was shown to exhibit a comparatively low rate of COD adsorption. Its saturation value was also below that of the Pittsburgh SGL and Brand A carbons.

4. The Pittsburgh SGL carbon proved superior to Brand A carbon, both in adsorptive rate and capacity. No saturation values were reached for these carbons during the study period, however.

It appears that the use of the Pittsburgh SGL, 8 x 30-mesh carbon should be continued. Tests of other types and brands will continue in the search for a better or more economical carbon.

Experimental Results of Run 4

A large portion of the cost of the tertiary process at South Tahoe is involved in coagulation and filtration ahead of the carbon columns. Although these steps perform several functions, the cost of this treatment has been partially justified as a necessary pretreatment ahead of the carbon columns. To investigate the necessity, or desirability, of this pretreatment at South Tahoe, two downflow pilot carbon columns were constructed. Without prefiltration, it was known that frequent backwashing of the carbon beds would be necessary, so a downflow system was selected. Also, concurrent evaluations were made of the effects of longer contact times and efficiency differences between upflow and downflow columns.

PCC 11, the downflow column treating separation bed effluent, was operated in parallel with the upflow columns; PCC 10 was the downflow column treating secondary effluent. Data collected from the operation of these two columns are presented in Figures 52 through 55.

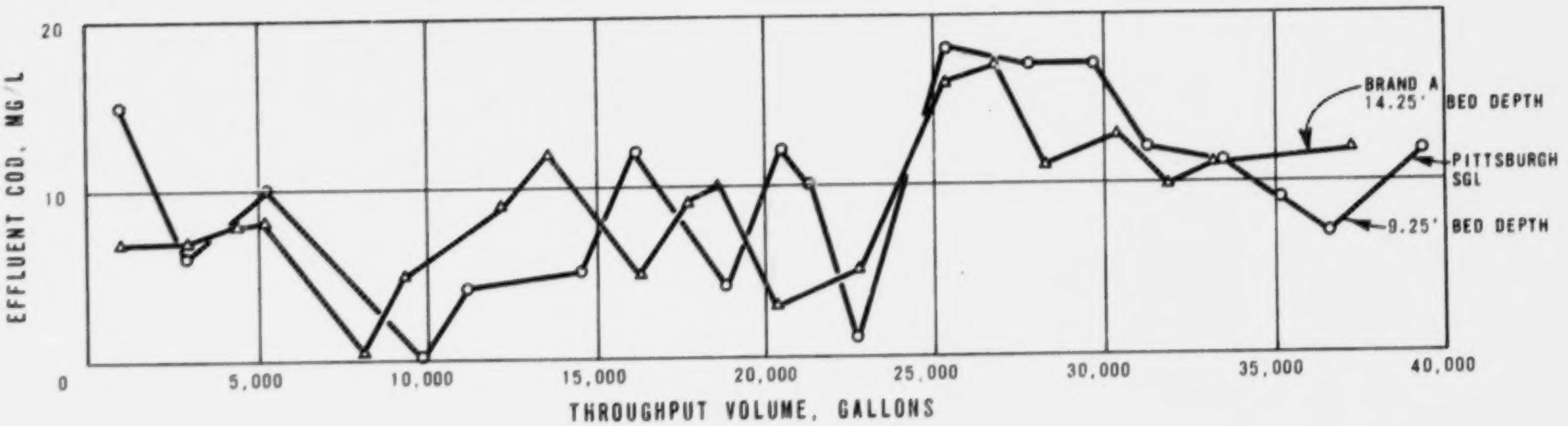


FIGURE 51

COMPARISON OF TWO CARBONS FOR COD REMOVAL

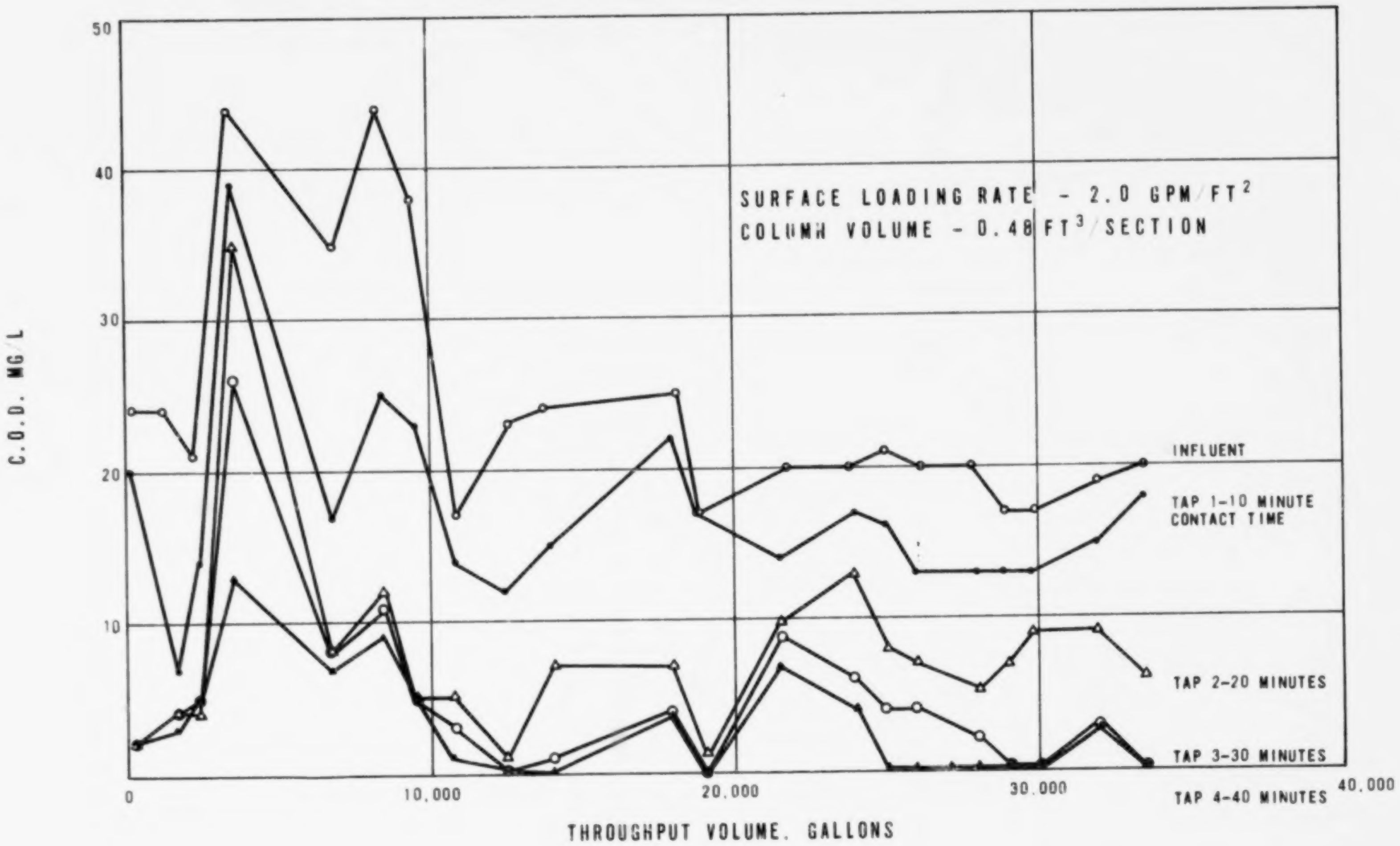


FIGURE 52

INFLUENT AND EFFLUENT C.O.D. CONCENTRATIONS FOR P.C.C. II
DOWNFLOW COLUMN TREATING SEPARATION BED EFFLUENT

16.

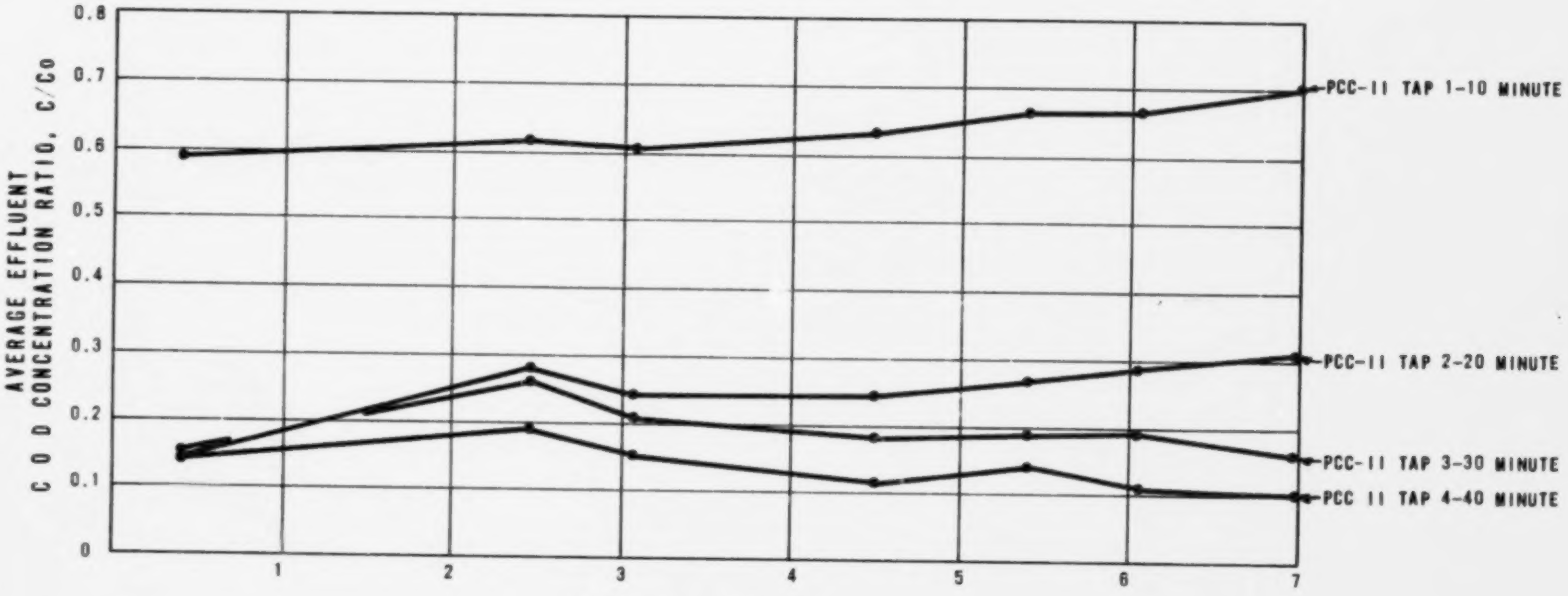


FIGURE 53

AVERAGE VALUES FOR THE RATIO OF EFFLUENT VERSUS INFLUENT C O D CONCENTRATION - P C C 11, DOWNFLOW COLUMN TREATING SEPARATION BED EFFLUENT.

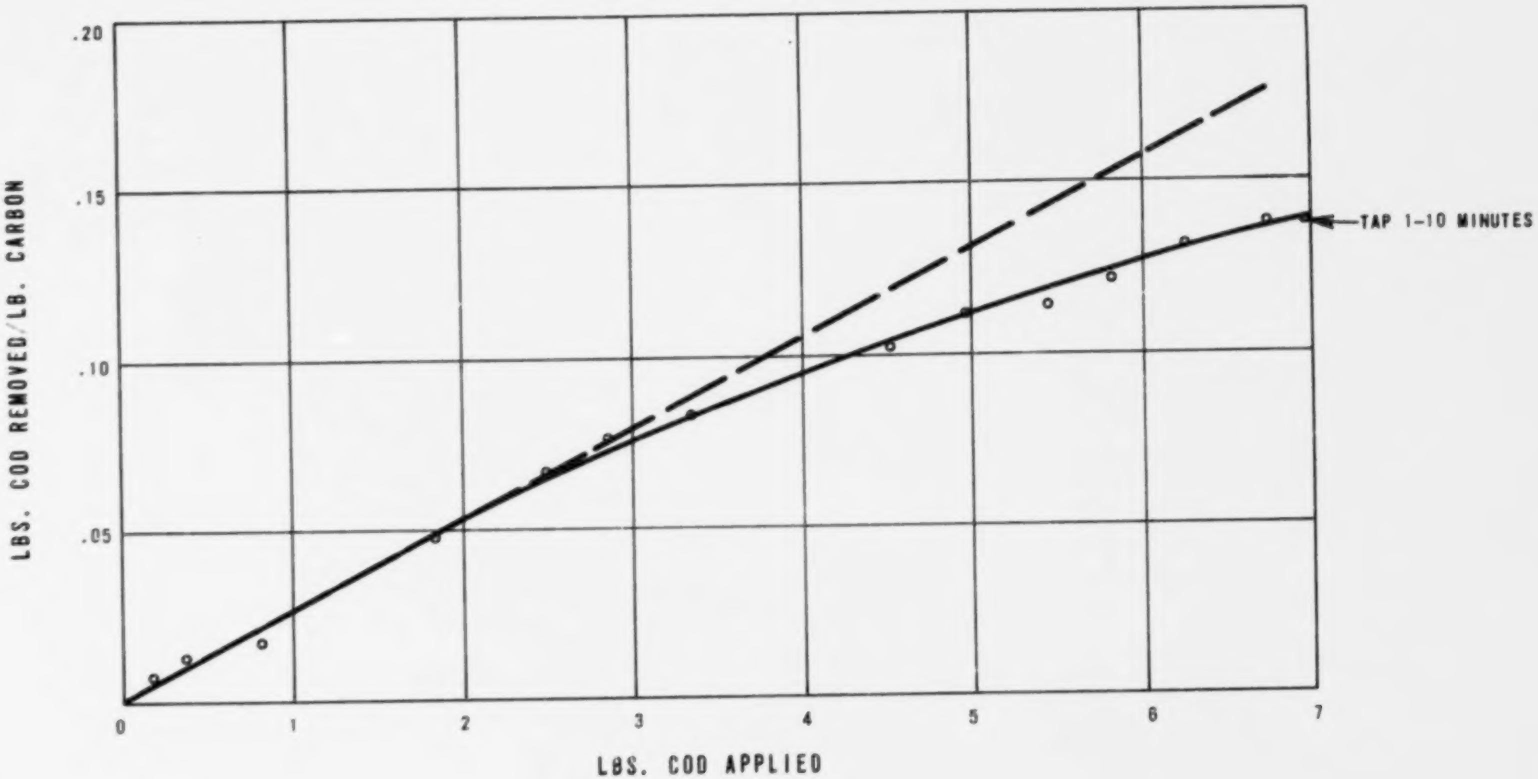


FIGURE 54

THE COD REMOVAL CAPACITY OF SECTION I OF PCC II
DOWNFLOW COLUMN TREATING SEPARATION BED EFFLUENT

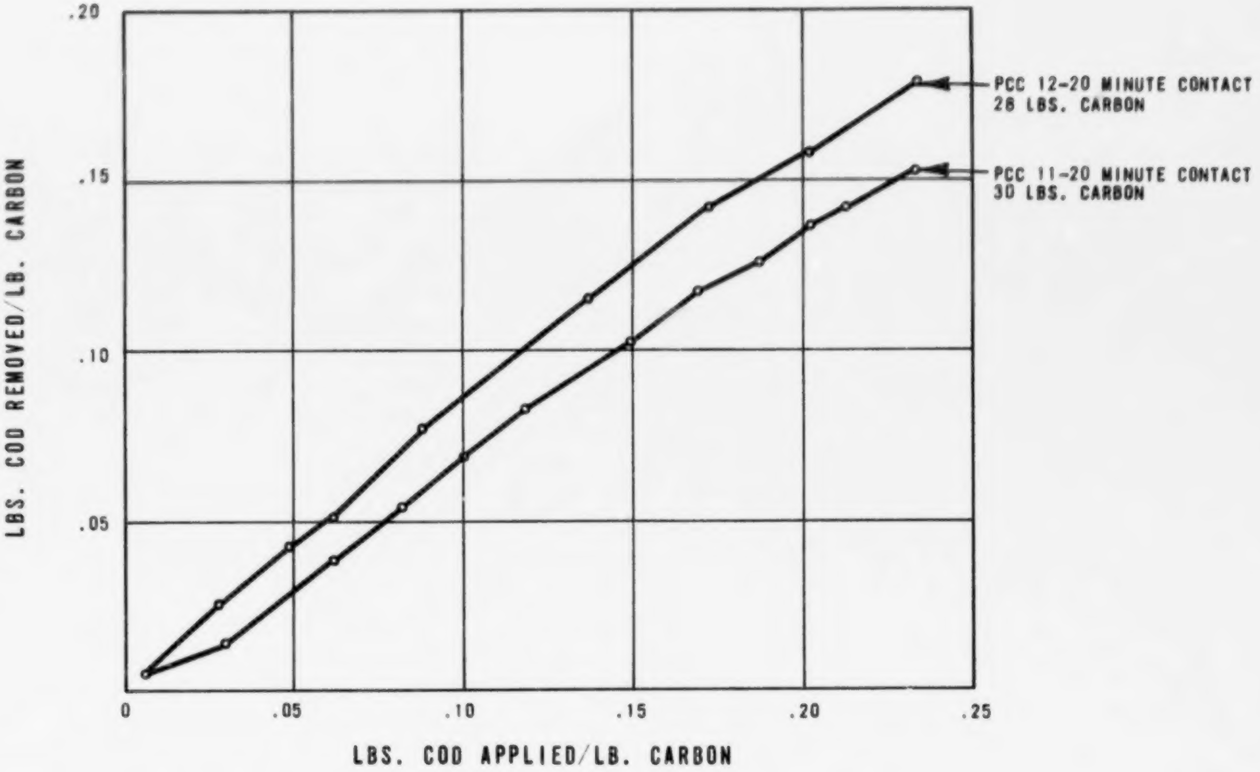


FIGURE 55

COMPARISON OF UPFLOW (PCC 12) AND DOWHFLOW (PCC 11) PILOT COLUMNS FOR THE REMOVAL OF COD

Effect of Contact Time. Figure 52 compares the influent COD concentration history to that of the effluents from taps 1, 2, 3, and 4 of PCC 11 (corresponding to 10, 20, 30, and 40 minutes contact time, respectively). It should be noted that the change in effluent COD per 10-minute increment of contact time was comparatively small after 20 minutes. This was especially true during the early part of the run, before the first column had become spent to an appreciable degree.

Figure 53 illustrates the effect of contact time somewhat better. Data taken from this plot at two points in time are summarized in Table 12. Note that after 3 pounds of COD had been applied, the change in C/C_o between taps 2 to 4 was minor. After 7 pounds of COD were applied, the incremental change in the average C/C_o between taps 2 and 3 had increased substantially. This would indicate that the effect of the exhaustion of a portion of the first section was beginning to show up as a higher COD concentration in the effluent from section 2. Figure 54 illustrates that after approximately 3 pounds of COD had been applied, the efficiency of the first section began to diminish, thus indicating noticeable exhaustion of a portion of the first section.

Table 12

VARIATION IN COD ADSORPTION WITH CONTACT TIME(1)

	3 lbs COD Applied				7 lbs COD Applied			
Tap	1	2	3	4	1	2	3	4
Contact Time, Minimum	10	20	30	40	10	20	30	40
Average C/C_o	0.61	0.25	0.21	0.15	0.70	0.33	0.16	0.10
Δ Average C/C_o		0.36	0.04	0.06		0.37	0.17	0.06

---

(1) PCC 11 — Downflow column treating separation bed effluent.

Upflow vs. Downflow Contactors. The above information confirms the desirability of an upflow, countercurrent contactor. Where a series of multiple-stage downflow contactors is used, the entire contents of one contactor logically must be regenerated at any particular time. Thus, if effluent quality is to be maintained, a large number of series contactors must be used if all carbon in the lead section is to be completely exhausted. With an upflow countercurrent contactor, any desired amount of carbon can be withdrawn, thus ensuring the withdrawal of only saturated carbon.

Figure 55 compares the COD removal efficiencies of the parallel-operated upflow and downflow columns, both at an approximate contact time of 20 minutes. The upflow column appeared to achieve better COD removal. Although several hypotheses can be made, the reason for this is not definitely known. The only difference between the columns was the surface loading rate, this being approximately 2 gpm/sq ft for the downflow column and 6 gpm/sq ft for the upflow column.

Effect of Pretreatment. COD concentration data compiled for PCC 10, the downflow column treating secondary effluent, are presented in Figure 56. Operational difficulties due to high solids carryover from the secondary clarifier were a problem from the start. During the periods of excessive solids carryover, the column could not be operated at all — the sludge would rapidly blind the carbon column to the point that no water would pass. Therefore, less than 15,000 gallons of secondary effluent were treated. The success of applying secondary effluent directly to granular carbon was thus shown to be highly dependent on the quality of secondary effluent. In a full-scale plant without filtration ahead of carbon columns, serious interruptions of plant operation could be expected from upsets in secondary treatment.

The extreme variation in the COD of the secondary effluent was dependent mostly on the suspended solids. After 7,000 gallons had been treated and a comparison between PCC 10 and PCC 11 was attempted, it became apparent that filtered COD tests on PCC 10 samples would be necessary. From that point on, all PCC 10 samples were analyzed, both unfiltered and after filtration through glass fiber filter paper.

The plots in Figure 56 indicate over 70 percent of the secondary effluent COD was in the form of suspended matter. A combination of filtration and adsorption in the carbon columns reduced the final effluent COD to an average of 16.3 mg/l. Over 50 percent of this total was due to solids penetrating the carbon columns. Also included in the figure for comparative purposes is the influent and effluent COD concentrations from PCC 11. The effluent from PCC 11 was of consistently better quality than the filtered effluent from PCC 10. This indicates the chemical coagulation was removing organic matter which cannot be removed by filtration alone.

Although some data were collected, a comprehensive study of the effects of pretreatment could not be accomplished due to the poor quality of the secondary effluent. Under the conditions experienced, pretreatment was mandatory for dependability.

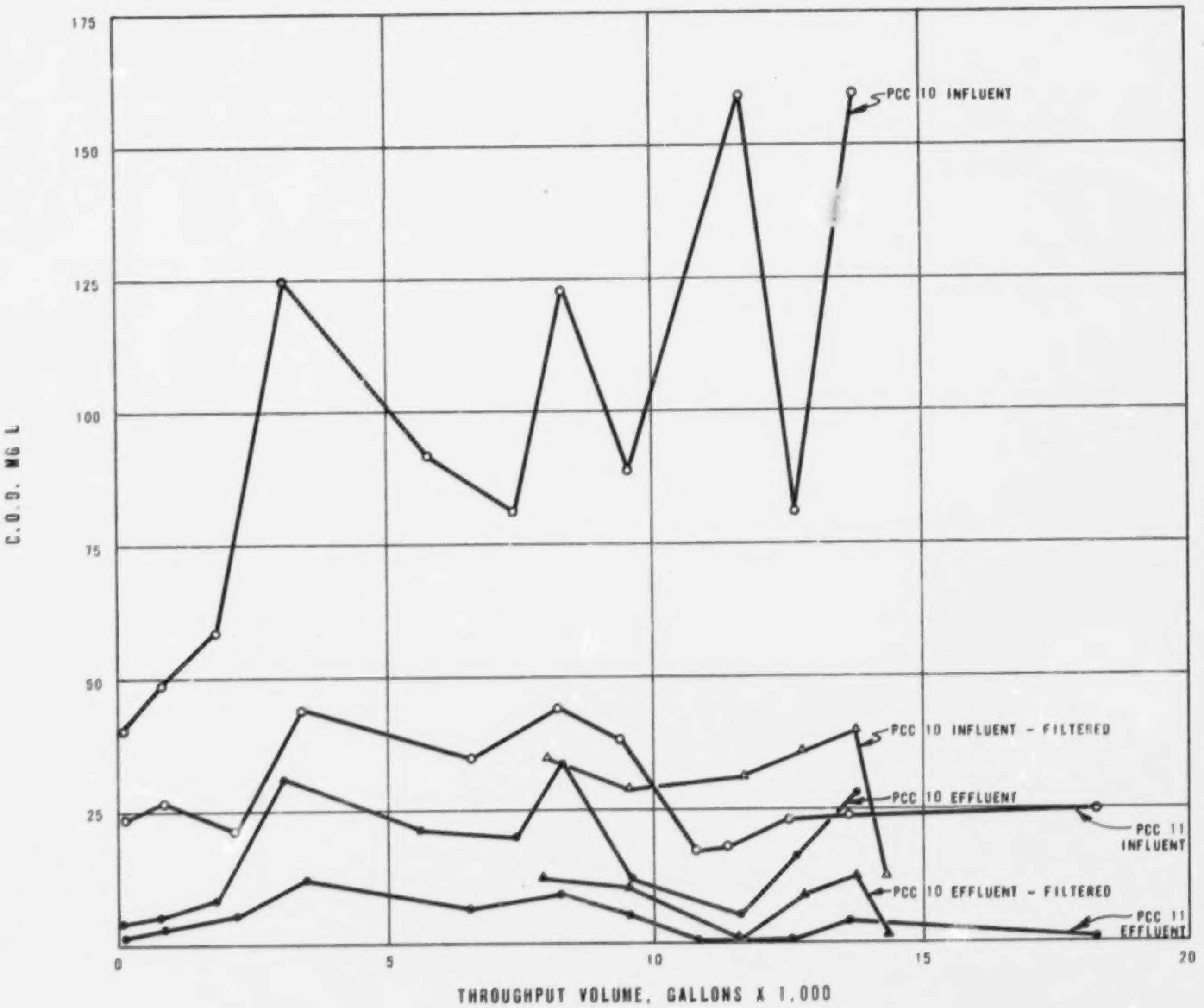


FIGURE 56

EFFECT OF PRETREATMENT ON COD REMOVAL

CHAPTER VI

RECOVERY AND REUSE OF COAGULANTS IN TREATING WASTE WATERS

GENERAL

The discharge of phosphorus-containing waste waters is recognized as an undesirable source of biological nutrients in many receiving bodies of water. Sawyer(18) and Lea, et al(19) have demonstrated that it is possible to remove the phosphates present in sewage by coagulation with alum.

As previously mentioned, over 98 percent of the phosphate in the secondary effluent is now being removed by the tertiary plant at the South Tahoe P. U. D. This installation is presently utilizing alum for the removal of phosphates. Since the alum constitutes 29 percent of the total cost of tertiary treatment (Table 7), this study was proposed to determine the economic feasibility of recovering and reusing chemicals at the South Tahoe P. U. D. In addition to reducing operating costs, chemical recovery would eliminate a difficult and expensive problem of sludge disposal.

ALKALINE METHOD OF ALUM RECOVERY

Literature Review

Lea, et al(19) suggested an alkaline method for alum recovery. The aluminum hydroxide on which the phosphates have been adsorbed is dissolved by raising the pH of the alum sludge to 11.9 with sodium hydroxide. The reaction converts the aluminum hydroxide to sodium aluminate and also returns the adsorbed phosphates into solution. Calcium chloride is then added to react with the phosphates to form insoluble calcium phosphate. The reactions for this process are as follows:(19)

$$\mathrm{Al(OH)_3 \cdot [PO_4]^{\equiv} + 4NaOH \xrightarrow{pH\ =\ 11.9} NaAlO_2 + Na_3PO_4 + 2H_2O + 3\,OH^-}$$

$$\mathrm{NaAlO_2 + 2Na_3PO_4 + 3CaCl_2 \longrightarrow NaAlO_2 + Ca_3(PO_4)_2 \downarrow + 6\,NaCl}$$

Lea, et al,(19) in pilot plant studies found reclaimed sodium aluminate and fresh alum to be equally effective in removing phosphates from sewage.

Contrary to the findings of Lea, et al, Sawyer(18) found that sodium aluminate was unsuitable for phosphate removal on Boston sewage unless supplemental caustic or alum or sulfuric acid were added to respectively raise or lower the pH into a favorable range. Sawyer also found that in the pH range of 10 to 11, considerably more sodium aluminate was required for phosphate removal than was required when alum alone was used. In addition, Sawyer found that sodium aluminate was not effective in removing the complex phosphates at high pH levels.

Packham,(20) studying coagulation with aluminum sulfate and sodium aluminate, found that sodium aluminate is only 25 percent as efficient as aluminum sulfate in the coagulation process. He proposed that this was due to the fact that sodium aluminate causes the pH to rise at the moment of coagulant addition, whereas alum causes the pH to drop at the moment of coagulant addition.

Sodium Aluminate as a Coagulant

Initial studies were performed to determine if sodium aluminate is a suitable coagulant for turbidity and phosphate removal on secondary sewage effluent at the South Tahoe P. U. D. Jar tests were performed on 500 ml samples of secondary effluent on a 6-paddle laboratory gang mixer. Liquid alum and reagent grade sodium aluminate were used in varying amounts. All percentages of alum and sodium aluminate are based on their equivalent contribution of aluminum ion, and all dosages are expressed as equivalent dry alum having the formula $Al_2(SO_4)_3 \cdot 14.3\ H_2O$. In addition, all samples were treated with 1 mg/l Dow Purifloc NP11, a non-ionic polyelectrolyte. After the samples were dosed with alum and sodium aluminate, in that order, and the polyelectrolyte added, all samples were flash-mixed for 30 seconds and then flocculated for 5 minutes. The samples were allowed to settle for a period of 1 hour, and then the clear liquor was analyzed for pH, turbidity, and phosphate. The results of these jar tests are tabulated in Table 13.

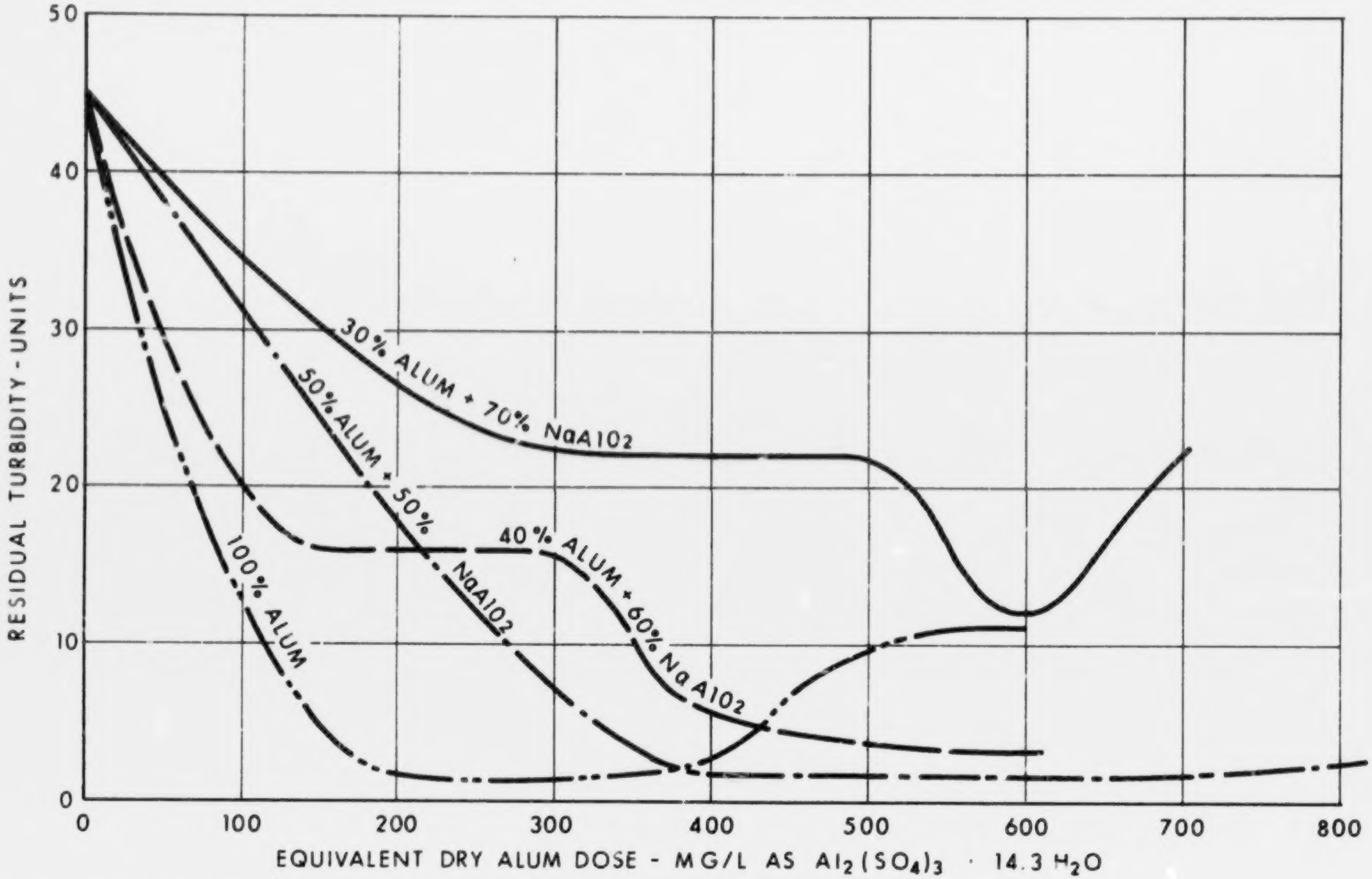
Figure 57 is a plot of residual turbidity vs. equivalent dry alum dose for various percentages of alum and sodium aluminate. A good residual turbidity can be obtained with 100 percent alum at a dose of 200 mg/l. As the percentage of sodium aluminate is increased, the residual turbidity at an equivalent alum dose of 200 mg/l is considerably higher than with alum alone. The equivalent alum dose had to be increased to 400 mg/l, using 50 percent alum and 50 percent sodium aluminate to obtain the same residual turbidity as that obtained with 100 percent alum at a dose of 200 mg/l. When 30 percent alum and 70 percent sodium aluminate were used, it was impossible to obtain good residual turbidity for the range of equivalent dry alum doses studied (200—700 mg/l). Jar tests were performed with 100 percent sodium aluminate, and no significant degree of coagulation was observed.

Figure 58 shows that as the percentage of sodium aluminate is increased for any equivalent dry alum dose, an increase in final pH is observed. Sodium aluminate is alkaline and consequently increases the alkalinity and pH of the

Table 13

JAR TESTS USING VARIOUS PERCENTAGES OF ALUM AND SODIUM ALUMINATE

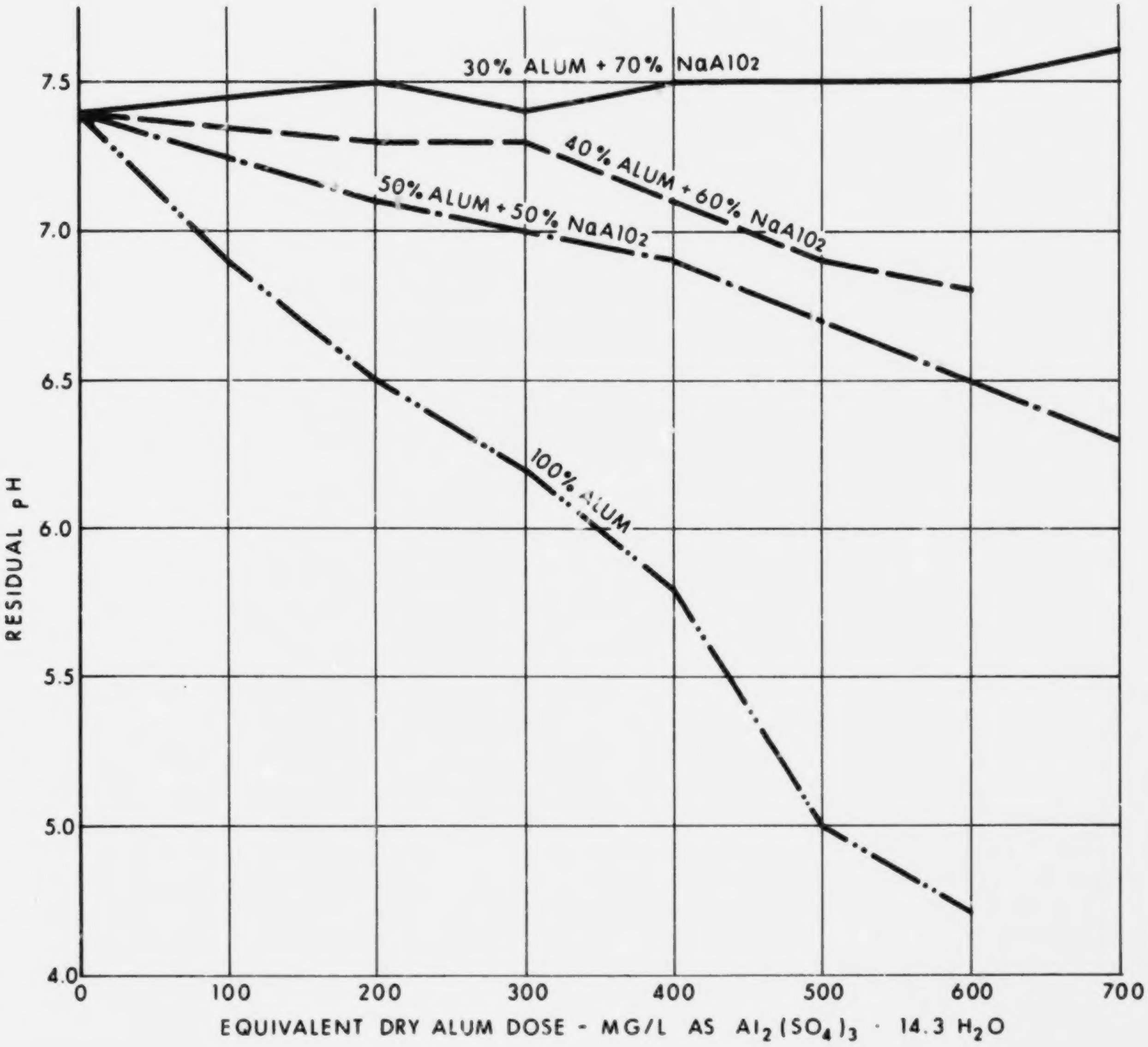
	Equivalent Dry Alum Dose (mg/l)	Turbidity	pH	Phosphate mg/l as PO_4^{---}
Secondary Effluent		45	7.4	44
100 Percent Alum	100	13	6.9	28
	200	1.5	6.5	2
	300	1.5	6.2	1
	400	2.1	5.8	0.6
	500	9.5	5.0	0.8
	600	11.0	4.7	2.3
50 Percent Alum, Plus 50 Percent Sodium Aluminate	200	18	7.1	23
	300	7	7.0	5
	400	1.9	6.9	4
	500	1.9	6.7	–
	600	1.7	6.5	–
	800	2.2	6.3	–
40 Percent Alum, Plus 60 Percent Sodium Aluminate	150	16	7.3	26
	200	16	7.3	28
	300	16	7.3	16.8
	400	5	7.1	2.8
	500	4.2	6.9	1.0
	600	3.1	6.8	0.6
30 Percent Alum, Plus 70 Percent Sodium Aluminate	200	27	7.5	26
	300	22	7.4	24.4
	400	22	7.5	20
	500	22	7.5	14.8
	600	12	7.5	3.2
	700	22	7.6	6.0



INITIAL TURBIDITY = 45 UNITS

FIGURE 57

RESIDUAL TURBIDITY VS. EQUIVALENT DRY ALUM DOSE
FOR
VARIOUS PERCENTAGES OF ALUM AND SODIUM ALUMINATE



INITIAL pH = 7.4

FIGURE 58

pH VS. EQUIVALENT DRY ALUM DOSE
FOR
VARIOUS PERCENTAGES OF ALUM AND SODIUM ALUMINATE

system. With this particular waste, the overall effect is to retard coagulation. Conversely, alum is acidic and tends to improve coagulation.

Figure 59 is a plot of residual phosphate concentration vs. equivalent dry alum dose. As was the case with turbidity, the equivalent alum dose has to be increased for increasing percentages of sodium aluminate in order to obtain the same phosphate removal as that obtained with alum alone. When 30 percent alum and 70 percent sodium aluminate is used, it was impossible to reduce the residual phosphate concentration below 3 mg/l. Figure 60 indicates that pH in the range of 5.0 to 7.0 is not a critical factor in optimum phosphate removal, as residual phosphate concentrations of less than 1 mg/l were obtained throughout this range.

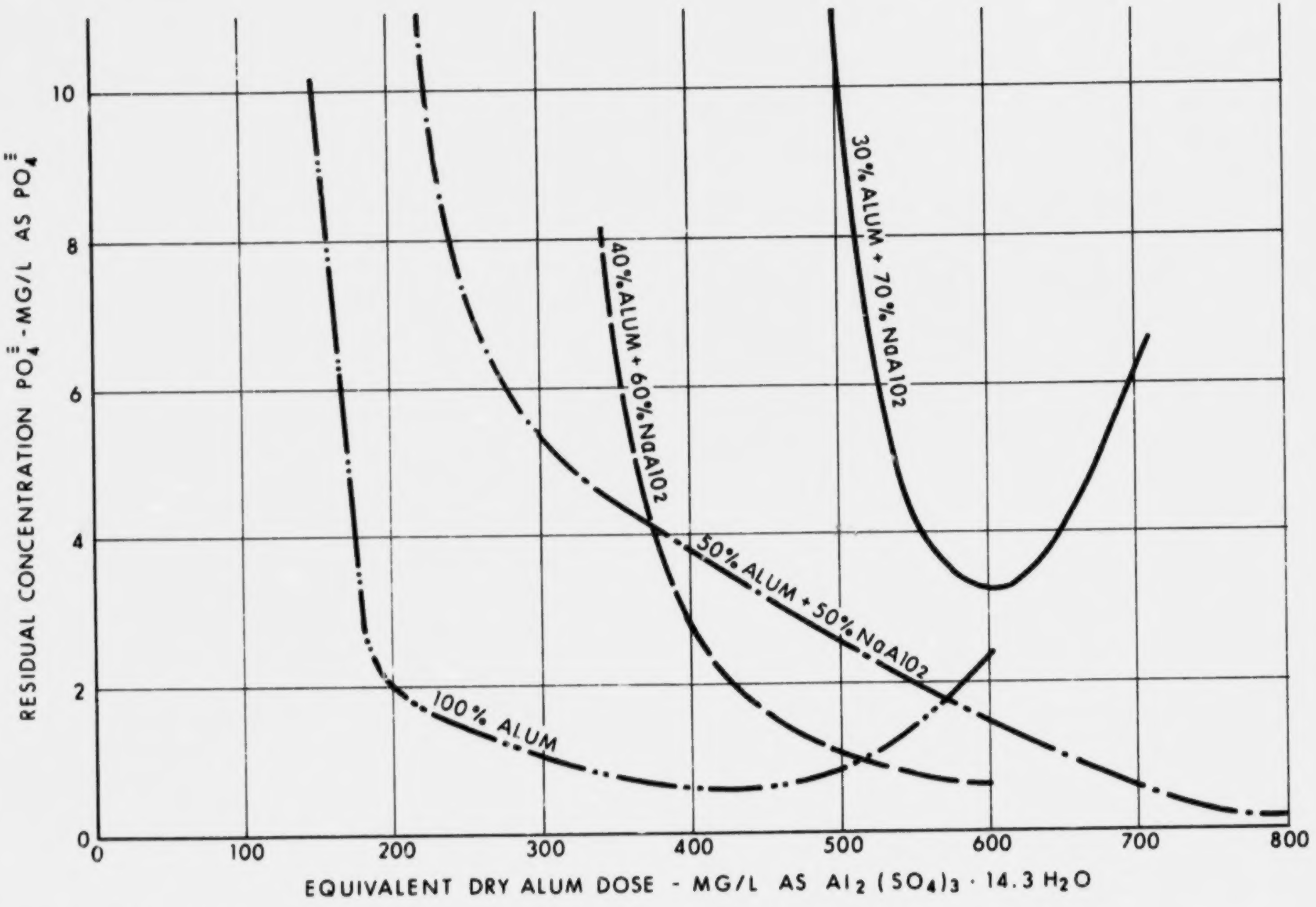
Alum Recovery by Sodium Hydroxide and Calcium Chloride

The initial studies of alum recovery consisted of batch tests in the laboratory to determine optimum conditions for alum recovery, preliminary chemical cost data, and problems which might be encountered in a plant-scale operation.

Alum sludge was obtained by adding 200 mg/l liquid alum and 1 mg/l Purifloc to 4 gallons of secondary effluent. The sewage was flash-mixed for 30 seconds and then flocculated for 5 minutes. The alum sludge was permitted to settle, and the clear liquor was decanted. The sludge was transferred to a smaller container and permitted to concentrate further by sedimentation.

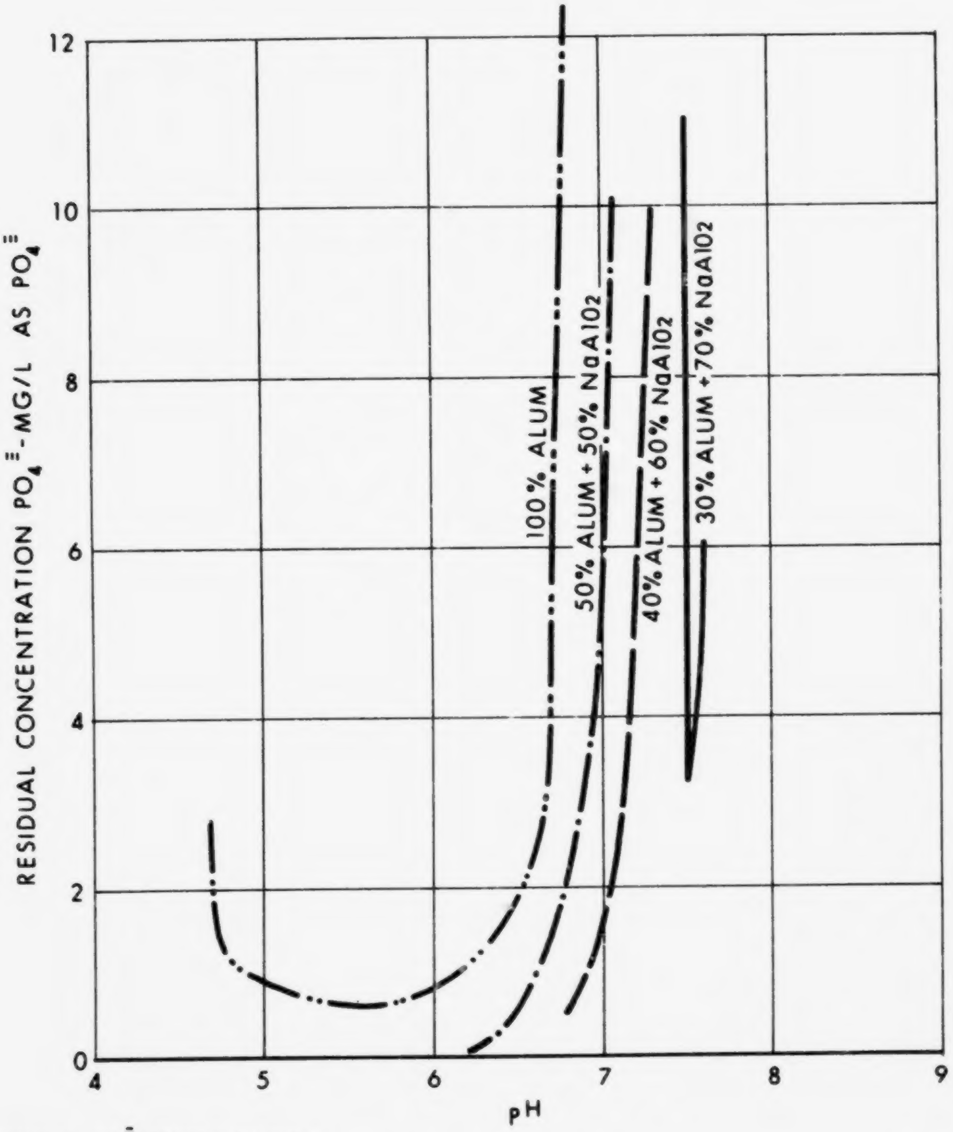
The resulting sludge was divided into 200 ml samples. The sludge samples were treated with varying quantities of sodium hydroxide to obtain pH values ranging from 10 to 12.5. After the sludge was dissolved, the aluminum and phosphate concentrations in the resulting solution were determined. The stoichiometric quantity of calcium chloride required to precipitate the phosphate as calcium phosphate was determined and added to the sodium aluminate solutions. The aluminum and phosphate concentrations in the final reclaimed alum solutions were determined. The volume reclaimed was determined by separation of the reclaimed alum from the calcium phosphate precipitate after 2 hours of sedimentation and filtration through a laboratory vacuum filter.

Figure 61 shows a comparison of the percent alum recovery obtained by filtration and sedimentation and the recoveries obtained by Lea & Rohlich.(19) At a pH of 11.9, Lea & Rohlich were able to recover 93 percent of the feed alum, whereas at Tahoe, only 75 percent recovery by sedimentation and 85 percent recovery by filtration were possible. About 10-percent increase in the alum recovery was achieved by filtering the resulting calcium phosphate sludge, rather than using the aluminate solution obtained by plain sedimentation.



INITIAL $PO_4^{\equiv}$ = 44 MG/L

FIGURE 59
RESIDUAL PHOSPHATE CONCENTRATION VS. EQUIVALENT DRY ALUM DOSE
FOR
VARIOUS PERCENTAGES OF ALUM AND SODIUM ALUMINATE



INITIAL $PO_4^{\equiv}$ CONC. = 44 MG/L

INITIAL pH = 7.4

FIGURE 60

RESIDUAL PHOSPHATE CONCENTRATION VS. pH
FOR
VARIOUS PERCENTAGES OF ALUM AND SODIUM ALUMINATE

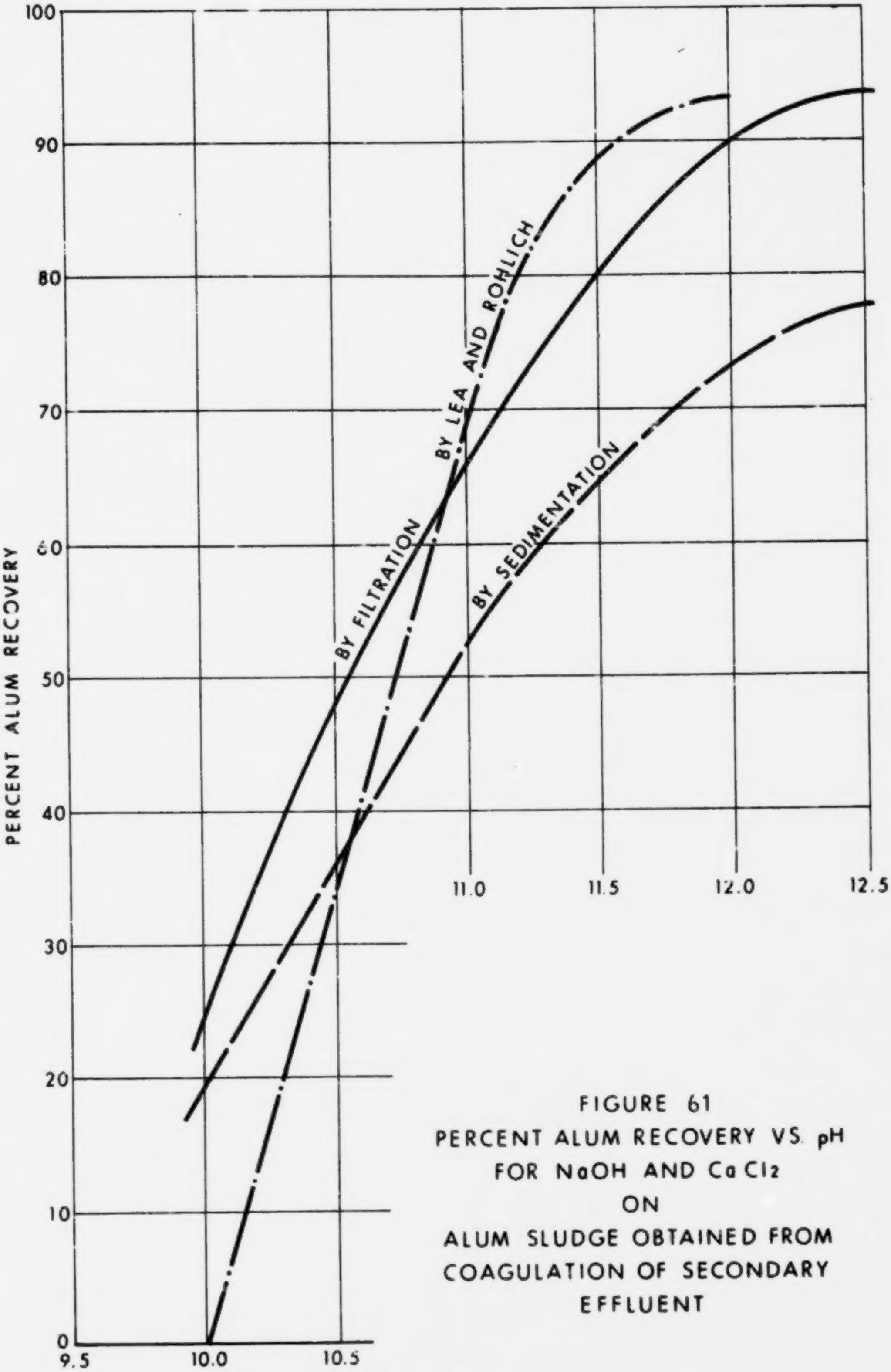


FIGURE 61
PERCENT ALUM RECOVERY VS. pH
FOR NaOH AND $CaCl_2$
ON
ALUM SLUDGE OBTAINED FROM
COAGULATION OF SECONDARY
EFFLUENT

Lea and Rohlich(19) state that aluminum hydroxide is converted to soluble sodium aluminate at pH 11.9. However, theory(21) indicates that aluminum hydroxide will dissolve and be converted to sodium aluminate between pH 10 and 10.5. Figure 62 shows that all the aluminum hydroxide is dissolved at pH 10 prior to the addition of calcium chloride. After the addition of calcium chloride, the final aluminum concentrations were drastically reduced between pH 10 and 11.5.

During the laboratory studies, an occasional side reaction was observed when the calcium chloride was added to the sodium aluminate solution. Following the addition of calcium chloride, the entire solution would occasionally turn into a white, gelatinous mixture. This resulting mixture would not dewater by either sedimentation or vacuum filtration. When this reaction occurred, the percent alum recovery was less than 5 percent. It was initially believed that the calcium chloride solution was too strong, causing a localized formation of calcium aluminate in the sodium aluminate mixture. However, this reaction was observed to occur when the strength of the calcium chloride solution was reduced to 0.1N.

Obviously, a side reaction occurs in this process, causing a large loss of aluminum in the pH range of 10 to 12. While no quantitative or qualitative data were obtained, it appears possible that the precipitation of insoluble calcium aluminate might be occurring. The pH dependence of this phenomena is not understood at this time.

Jar tests were performed on samples of secondary effluent dosed with various percentages of makeup alum, and alum recovered at pH 11.9 with the sodium hydroxide-calcium chloride method. The initial tests consisted of dosing with 100 percent reclaimed alum. Since no significant coagulation was obtained, the percent of reclaimed alum added was continually reduced until coagulation was obtained. Coagulation was obtained at 50 percent makeup alum and 50 percent reclaimed alum, using an equivalent dry alum dose of 400 mg/l, but the residual turbidity was reduced to only 20 units.

Jar tests were performed adding sulfuric acid to the secondary effluent prior to the addition of the makeup alum and reclaimed alum. Excellent coagulation was obtained at 50 percent alum and 50 percent reclaimed alum, using an equivalent dry alum dose of 400 mg/l and 150 mg/l sulfuric acid.

It is recognized that an excess amount of alkalinity is present in the reclaimed alum solution due to the fact that the alum sludge was not thickened prior to the reclamation process. However, thickening the alum sludge increases both the complexity and capital cost of the process and was not believed to be justified at this time.

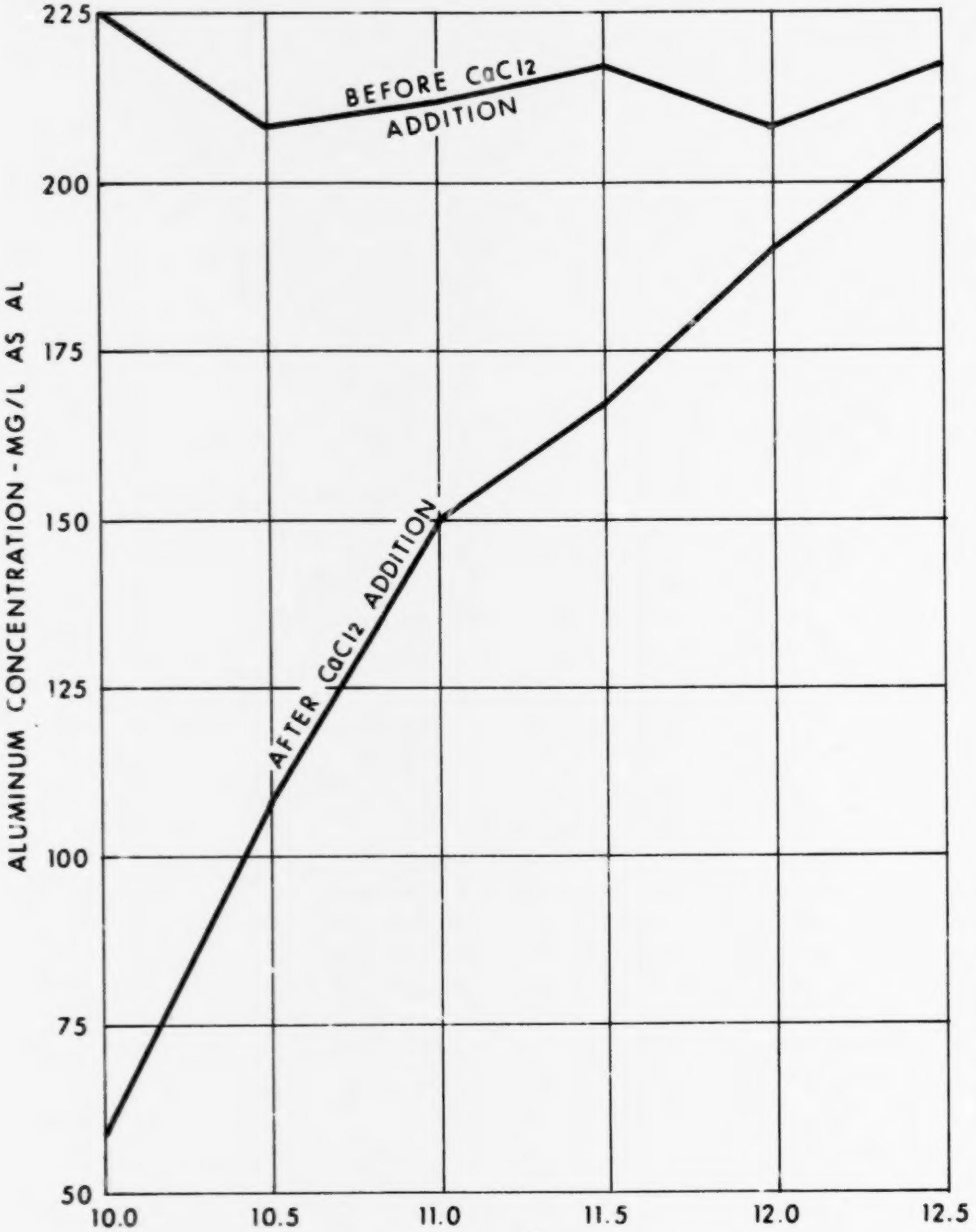


FIGURE 62

ALUMINUM CONCENTRATION VS. pH
BEFORE AND AFTER CALCIUM CHLORIDE ADDITION

The cost of chemicals used in the recovery methods studied is shown in Table 14. These costs are f.o.b., South Lake Tahoe, California.

Table 14

CHEMICAL COSTS
(f.o.b., South Lake Tahoe, California)

Chemical	$/ton
Sodium Hydroxide	72.50
Calcium Chloride	70.00
Lime as 100% $Ca(OH)_2$	28.00
Sulfuric Acid	34.25
Liquid Alum (dry basis)	57.40

Table 15-1 shows the chemical cost to reclaim and make up one ton of equivalent dry alum, using sodium hydroxide and calcium chloride. The lowest chemical cost is obtained at pH 11.5, where the cost of coagulant is $69.90 per ton by sedimentation and $62.20 per ton by filtration. Since fresh alum costs $57.40 per ton, it is evident that it is not economical to reclaim alum by this method. The total cost to reclaim alum will actually be higher than that shown in Table 15-1, because the capital, operational, and power costs have not been included.

Alum Recovery by Lime

Since the scheme to recover alum by using sodium hydroxide and calcium chloride did not appear practical, alum recovery using lime was attempted. Lime contains both the hydroxyl and calcium ions required to dissolve the aluminum hydroxide and precipitate the phosphate. There are two apparent advantages in using lime rather than sodium hydroxide and calcium chloride. First, lime costs $28 per ton, whereas sodium hydroxide and calcium chloride each cost about $70 per ton. Second, this scheme would require the handling of only one chemical rather than two chemicals.

Samples of alum sludge from secondary effluent were obtained as previously described. Alum sludge samples (200 ml) were treated with varying amounts of reagent grade calcium hydroxide. After mixing and allowing to settle for 2 hours, the volume of reclaimed alum obtained by sedimentation and filtration was determined. Analyses for pH, phosphate, and aluminum were performed on the relcaimed alum solution.

Figure 63 shows percent alum recovery vs. pH with lime as a reactant. The maximum recovery was obtained at pH of 10.1. At this pH, the recovery by sedimentation was 19.7 percent, and by filtration, 35.2 percent. Table 15-2

Table 15-1

CHEMICAL COST TO RECLAIM AND MAKE UP ONE TON EQUIVALENT DRY ALUM USING SODIUM HYDROXIDE AND CALCIUM CHLORIDE

	Sedimentation				Filtration			
pH	Percent Recovery	Cost Reclaimed Alum	Cost Makeup Alum	Total Cost $/Ton	Percent Recovery	Cost Reclaimed Alum	Cost Makeup Alum	Total Cost $/Ton
10.0	19.1	$34.70	$46.40	$81.10	24.1	$24.40	$43.60	$68.00
10.5	36.2	40.40	36.60	77.00	49.6	40.60	29.00	69.60
11.0	52.7	44.60	27.20	71.80	68.0	44.60	18.40	63.00
11.5	62.5	48.40	21.50	69.90	75.6	48.20	14.00	62.20
12.0	74.6	55.50	14.60	70.10	84.9	55.90	8.70	64.60
12.5	77.5	83.30	12.90	96.20	97.4	83.10	1.50	84.60

Table 15-2

CHEMICAL COST TO RECLAIM AND MAKE UP ONE TON EQUIVALENT DRY ALUM USING LIME

	Sedimentation				Filtration			
pH	Percent Recovery	Cost Reclaimed Alum	Cost Makeup Alum	Total Cost $/Ton	Percent Recovery	Cost Reclaimed Alum	Cost Makeup Alum	Total Cost $/Ton
8.4	6.8	$ 3.60	$53.50	$57.10	13.4	$ 3.60	$49.70	$53.30
9.0	5.2	5.40	54.40	59.80	11.5	5.30	50.80	56.10
9.3	9.6	7.20	51.80	59.00	18.2	7.15	47.00	54.15
10.1	19.7	11.10	45.10	56.20	35.2	10.65	37.20	47.85
11.3	14.5	14.25	49.10	63.35	20.6	14.10	45.60	59.70

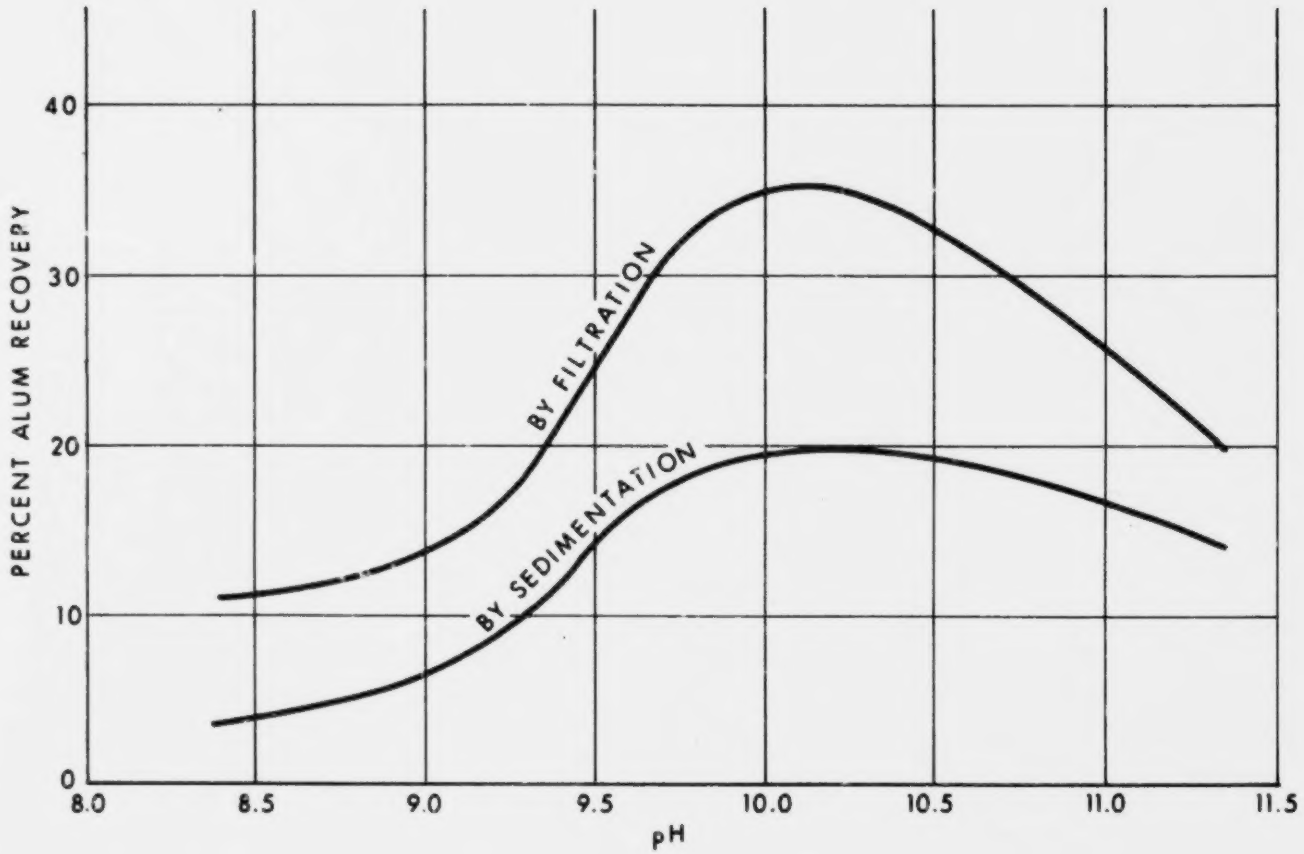


FIGURE 63
PERCENT ALUM RECOVERY VS. pH FOR LIME
ON
ALUM SLUDGE SAMPLES FROM COAGULATION OF SECONDARY EFFLUENT

shows the chemical cost required to reclaim and make up one ton of equivalent dry alum for the various pH's studied. At a pH of 10.1, the chemical cost is $47.85 per ton by filtration and $56.20 per ton by sedimentation.

While the percent recovery using lime is considerably less than using sodium hydroxide and calcium chloride, the cost of chemicals is approximately $18 per ton of alum less when using lime. In addition, the white gelatinous substance occasionally noted in the studies of the sodium hydroxide-calcium chloride method did not form when lime was used. This would reduce the possibility of losing an entire batch of alum sludge, plus the chemicals used for alum recovery. The resulting calcium phosphate sludge dewatered much more readily than did the final sludge obtained by using sodium hydroxide and calcium chloride.

Since the data indicated that a savings in chemical cost might be realized if alum recovery with lime were practiced, an estimate of the capital and operating costs was made. It is estimated that an alum recovery plant for a 5-mgd plant would cost $60,000. Amortized for 15 years at 4 percent interest, this is equivalent to a daily cost of $14.80. It is estimated that 8 hours of operation will be required per day at $3.75 per hour, which would result in a daily operating cost of $30. The total daily cost is $44.80.

A 5-mgd plant receiving an alum dose of 200 ppm will require 4.17 ton of alum per day. The studies indicate that 35 percent alum recovery can be obtained by filtration at pH 10.1. Therefore, 1.46 tons of alum would be reclaimed daily. Thus, the capital and operating cost would be $44.80/1.46, or $30.60 per ton reclaimed alum.

Table 15-2 shows that the chemical cost is $10.65/.352, or $30.30 per ton reclaimed alum. Therefore, the total cost to reclaim one ton of alum is $60.90. Since fresh alum costs $57.40 per ton, it is apparent that it is uneconomical to reclaim alum by the lime method of recovery.

ACID METHOD OF ALUM RECOVERY

Literature Review

Since the alkaline method of alum recovery appeared to be uneconomical, an acid recovery method was attempted.

Roberts and Roddy(22) reclaimed and reused alum sludge from a domestic water treatment plant at Tampa, Florida, using an acid recovery scheme. The alum sludge was pumped from the clarifiers to a basin, where the sludge was allowed to thicken. The thickened sludge then went to a reactor where concentrated sulfuric acid was added. The resulting alum solution was then used as a coagulant for the raw water.

Experience at Tampa showed that the pH of the resulting aluminum sulfate solution ranged from 1.5 to 2.5, depending upon the alkalinity of the raw water being treated. While they did not discuss the effects of the dissolved color in the reclaimed alum solution, there was no change in the chemical or bacteriological quality of the treated water while using the reclaimed alum.

Gauntlett(23) reports the following studies of acid recovery of alum:

1. Palin(24) reclaimed alum by charring the sludge at 400°C and then dissolving the burned material in sulfuric acid. He found that the volume of insoluble material remaining for disposal was not significantly reduced by acid treatment. Palin concluded that the cost to reclaim alum was too high because of the cost of the chlorine required to remove the color from the reclaimed alum solution.

2. Isaac and Vahidi(25) were able to recover two thirds of the alum at pH 3. The percent alum recovery increased slightly when the pH was reduced below 3. Contrary to the findings at Tampa,(22) they found that the dissolved color reduced the effectiveness of reclaimed alum.

Alum Recovery by Sulfuric Acid

In the studies conducted at South Tahoe, alum sludge was obtained from coagulation of secondary effluent. Samples of this sludge were dosed with sulfuric acid to obtain final pH's ranging from 1 to 3. The aluminum and phosphate concentrations, along with the volumes of reclaimed alum obtained by filtration and sedimentation were determined. These results are shown in Table 16.

Table 16

ALUM RECOVERY USING SULFURIC ACID

100 ml alum sludge samples, obtained from coagulation of secondary effluent, containing 31.6 mg aluminum.

pH	mg H_2SO_4 Required	Phosphate (mg/l)	Aluminum (mg/l)	Final Reclaimed Vol. – ml Sedimentation	Final Reclaimed Vol. – ml Filtration
3.0	63.8	70	150	55	88
2.5	137.4	550	331	84	90
2.0	236	570	331	87	93
1.5	451	570	335	90	95
1.0	1,582	570	380	98	103

Figure 64 shows the percent alum recovery vs. pH. The curve shows that essentially 100 percent recovery is possible at pH 2. From 95 to 100 percent recovery is possible at pH 2.5. Figure 65 shows that essentially all of the phosphate is dissolved at pH 2.5.

It is not surprising that Isaac and Vahidi(25) were able to recover only 67 percent alum at pH 3. Figures 64 and 65 indicate that the aluminum hydroxide is not completely dissolved at pH 3. Furthermore, theory(21) indicates that aluminum hydroxide completely dissolves between a pH of 2.5 and 3.

Table 17 shows the chemical costs required to reclaim and make up one ton of equivalent dry alum when phosphates are not removed from the reclaimed alum. At a pH of 2.5, the chemical cost of reclaimed alum is $20.30 per ton by sedimentation and $16.60 per ton by filtration, as compared to $57.40 per ton for fresh alum.

Phosphate Removal From Acidified Alum Sludge

Since phosphate removal is one of the purposes of the tertiary treatment plant at South Tahoe, a process had to be developed to remove the phosphates present in the alum solution reclaimed by the acid method. If the phosphates were not removed from the reclaimed alum solution, then phosphate would be recycled, and removal would not be accomplished in tertiary treatment process.

Two methods were studied to remove the phosphate from the alum reclaimed with sulfuric acid. These were ion exchange and activated alumina.

In reviewing the literature, no work was found with regard to removing phosphates from an acid alum solution by ion exchange. Without going into the details of ion exchange, it should be apparent that to achieve phosphate removal, an anionic ion exchange resin would be required.

No literature on phosphate removal by adsorption on activated alumina was available. However, phosphate is second only to hydroxyl ion in the usual adsorption series for aluminas.(26)

Ion Exchange. A 1-1/2-inch-diameter column was constructed and charged with an anionic resin (Duolite A-102D) to an effective bed depth of 34 inches. The resin was supported on a plug of glass wool, and the flow rate was adjusted by a valve at the base of the column.

Alum sludge, from coagulation of secondary effluent, was dosed to a pH of 2.5 with sulfuric acid and was passed through the resin on a chloride cycle. The acid-treated solution had an aluminum concentration of 300 mg/l and a phosphate concentration of 440 mg/l. Phosphate analyses were made for every 200 ml passing through the column. When the effluent phosphate

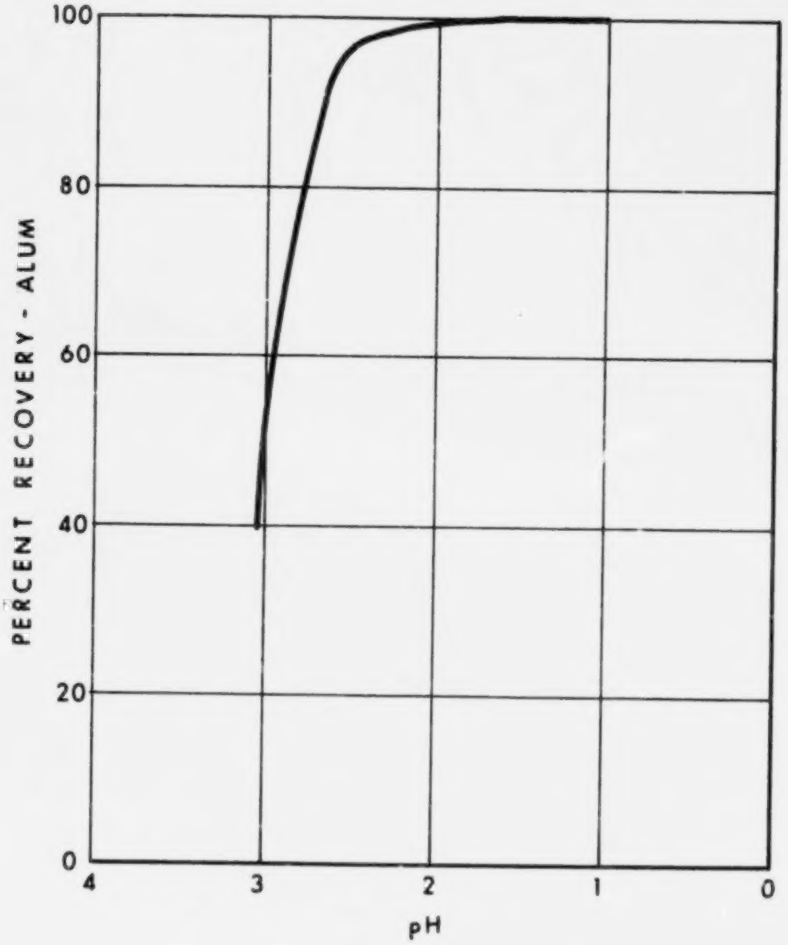


FIGURE 64
ACID RECOVERY OF ALUM
pH VS. PERCENT ALUM RECOVERY

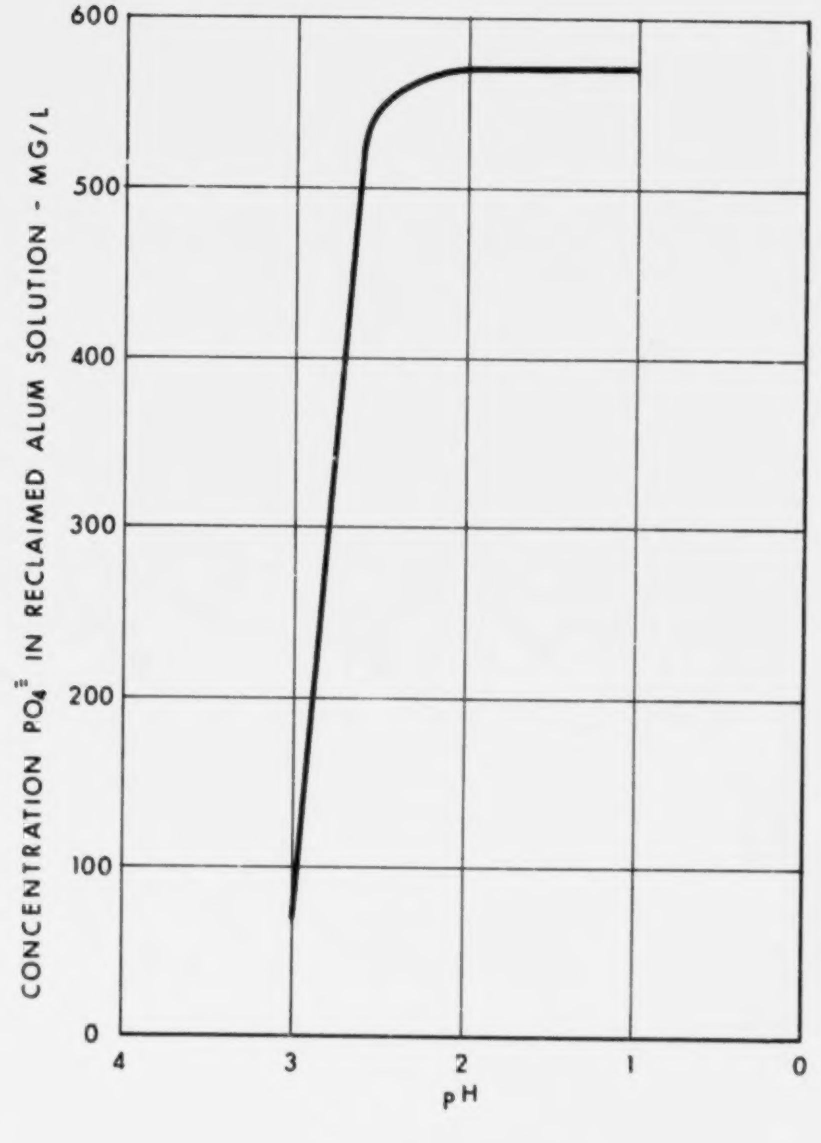


FIGURE 65

ACID RECOVERY OF ALUM

pH VS. CONCENTRATION $PO_4^{\equiv}$ IN RECLAIMED ACID SOLUTION

Table 17

CHEMICAL COST TO RECLAIM AND MAKE UP ONE TON EQUIVALENT DRY ALUM USING SULFURIC ACID

(No Phosphate Removal)

	Sedimentation				Filtration			
pH	Percent Recovery	Cost Reclaimed Alum	Cost Makeup Alum	Total Cost $/Ton	Percent Recovery	Cost Reclaimed Alum	Cost Makeup Alum	Total Cost $/Ton
3.0	26.2	$ 6.27	$42.40	$ 48.67	41.8	$ 6.21	$33.40	$ 39.61
2.5	88	13.41	6.92	20.32	94.4	13.38	3.22	16.60
2.0	91.2	23.10	5.06	28.16	97.5	23.00	1.43	24.43
1.5	95.6	44.20	2.53	46.73	100^+	43.80	—	43.80
1.0	100^+	154.10	—	154.10	100^+	154.10	—	154.10

concentration reached 100 mg/l, the run was terminated, and the resin was then regenerated with 2 bed volumes of 10 percent sodium chloride, followed by a rinse with 5 bed volumes of dionized water.

An average of three runs showed that 1.2 bed volumes of acid-treated alum sludge could be passed through the resin before the phosphate broke through. It was noted that the total phosphate removal decreased with each successive run. This reduction of exchange capacity was undoubtedly due to organic fouling of the resin by the feed solution, which had a COD of 200 mg/l. No reduction in the aluminum concentration was observed when the reclaimed alum was passed through the resin.

A preliminary estimate of the salt requirements for regenerating the resin was made. Using an effluent aluminum concentration of 300 mg/l, 1.2 bed volumes of treated liquor passing through the column between regenerations, and the previously mentioned regeneration quantities, 53.5 tons of salt would be required per ton of reclaimed alum. The price of salt is $25 per ton, f.o.b., South Lake Tahoe. Therefore, the cost of salt to reclaim one ton of relatively phosphate-free dry alum is $1,340. As a result of this high cost, the study of ion exchange for phosphate removal was abandoned.

Activated Alumina. A 1-1/2-inch-diameter column was constructed and charged with 1/4-inch x 8-mesh active alumina (Kaiser Chemical KA-101) to a depth of 30 inches. This alumina has an active surface area of 360 m^2/g. The alumina was supported on a plug of glass wool, and the flow rate was adjusted by a valve at the base of the column.

Alum sludge, from coagulation of secondary effluent, was adjusted to pH 2.5 with sulfuric acid. The resulting solution was passed through the column of activated alumina. The feed solution had an aluminum concentration of 250 mg/l and a phosphate concentration of 440 mg/l. When the effluent phosphate concentration reached 100 mg/l, the run was terminated and the aluminum was reactivated. Reactivation consisted of passing 5 bed volumes of 1N H_2SO_4 through the alumina column followed by a rinse of 2 bed volumes of deionized water. On the first run, using virgin alumina, 3.5 bed volumes were passed through the column before breakthrough occurred. Following reactivation with 1N H_2SO_4, it was impossible to get the effluent phosphate below 115 mg/l. On the third run, the final phosphate concentration could not be brought below 150 mg/l.

This loss of capacity was partially due to the fact that the alumina was being dissolved by the acid alum solution. While the influent aluminum concentration was 250 mg/l, the effluent aluminum concentration was 580 mg/l. In addition, the pH of the reclaimed alum solution was observed to rise from 2.5 to 3.5 when passing through the column. Since activated alumina is approximately 90 percent Al_2O_3, the following reaction was probably occurring in the column:

$$Al_2O_3 + 3\ H_2SO_4 \longrightarrow Al_2(SO_4)_3 + 3\ H_2O$$

A preliminary estimate of the sulfuric acid requirements for reactivating the alumina was made. Using an effluent aluminum concentration of 580 mg/l, the 3.5 bed volumes of alum solution passing through the column between reactivations and the previously mentioned regeneration quantities, 10.5 tons of sulfuric acid would be required to reclaim one ton of alum. The cost of sulfuric acid delivered to Lake Tahoe is $34.25 per ton. Therefore, the cost of chemicals to reactivate the alumina is $360 per ton of reclaimed alum, which is prohibitively high.

Other methods of reactivating the alumina were attempted. These included:

1. Heating the alumina at 450°F for 1 hour.
2. Washing with one volume of 1N H_2SO_4, followed by heating at 450°F for 1 hour.
3. Rinsing the alumina with two volumes of 1N NaOH, followed by two volumes of nitric acid.
4. Washing with five volumes of 1N H_2SO_4 (as control).

Twenty-five gram samples were reactivated as above. The samples were then placed into the acid alum solution. All jars contained the same quantity of phosphate. The alumina was agitated in the solutions for 1 hour. The final phosphate concentrations were determined. The results showed that the degree of removal was the same for all methods of regeneration. Since no increase in the adsorptive capacity was observed when using different methods of reactivation, the studies using activated alumina were abandoned.

RECOVERY AND REUSE OF LIME

Introduction

Since the recovery and reuse of alum is not economically feasible at the South Tahoe P. U. D., studies were performed on the feasibility of recovery and reuse of lime for coagulation of sewage. The practice of recovering lime by recalcining is not new. The pulp and paper industry has been using lime recalcining furnaces for many years. In a number of municipal water softening plants, lime is routinely reclaimed in recalcining operations. The recovered lime is reused many times, with little or no loss of efficiency in the water softening process.

While work was being performed on alum recovery, it was found that 98 percent of the ammonia-nitrogen present in the sewage at South Tahoe could be removed by ammonia stripping. Briefly, this process would consist of raising the pH of the sewage to 10.8 and passing it through a forced draft aeration tower. The use of lime would not only remove the phosphates present in the secondary effluent by the precipitation of calcium phosphate, but elevate the pH sufficiently so that ammonia stripping could be accomplished. The details of the ammonia stripping tower are discussed in Chapter VII.

Literature Review

When lime is added to sewage, it reacts with the alkalinity present in the water to form calcium carbonate and to release hydroxyl ions, which raise the pH of the water. Excess calcium ions react with the phosphates present in the sewage to form insoluble calcium phosphate. These reactions are shown in the following equations:

$$CaO + HCO_3^- \longrightarrow CaCO_3\downarrow + OH^-$$

$$\underset{\text{excess}}{3\,Ca^{++}} + 2\,PO_4^{\equiv} \longrightarrow Ca_3(PO_4)_2\downarrow$$

Sawyer(18) found that the total soluble phosphorus could be reduced to 0.5 mg/l with a lime dose of approximately 300 mg/l as $Ca(OH)_2$. Owen,(27) using the effluent from a low-rate trickling filter, reduced the soluble phosphorus concentration with 1 hour of settling to 1.3 mg/l, using 300 mg/l CaO and to 0.25 mg/l, using 600 mg/l CaO. When these samples were filtered, the residual phosphate concentrations were 0.6 and $<$0.01 mg/l, respectively. In addition, Owen(27) found that a pH of 11 was sufficient to precipitate the phosphorus as calcium phosphate.

Rand and Nemerow(28) found that it is possible to reclaim and reuse lime for the coagulation of raw sewage. Laboratory studies showed an average of 87 percent recovery of the feed lime by recalcining the resulting lime sludge. The lime sludge contained sufficient volatile material to support its own combustion without auxiliary fuel and to recalcine the lime. No change in phosphate removal was observed when the reclaimed lime was recycled with 13 percent makeup lime. The economic analysis presented indicates that lime recovery may be economically feasible.

Feasibility of Lime as a Coagulant

Jar tests using lime as a coagulant were performed on 500 ml samples of secondary effluent. Lime was added in varying amounts, and one mg/l Purifloc was added to all samples. Following chemical addition, the samples were flash-mixed for 30 seconds and flocculated for 5 minutes. After 1 hour of clarification, the clear liquor was analyzed for pH, phosphate, and turbidity.

Figure 66 shows that the phosphate can be reduced to 1.2 mg/l with 1 hour of clarification, using a lime dose of 400 mg/l as CaO. As shown on Figure 67, a lime dose of 400 mg/l corresponded to a pH of 11.6, which is more than adequate for ammonia stripping. Figure 68 shows a plot of residual turbidity vs. lime dose. The jar tests indicate that lime is a poor coagulant for turbidity removal. However, this residual turbidity was observed to be a fine suspension of lime, which could be removed by filtration. Filtration would not only remove the suspended material, but reduce the residual phosphate concentration, as verified under pilot plant studies.

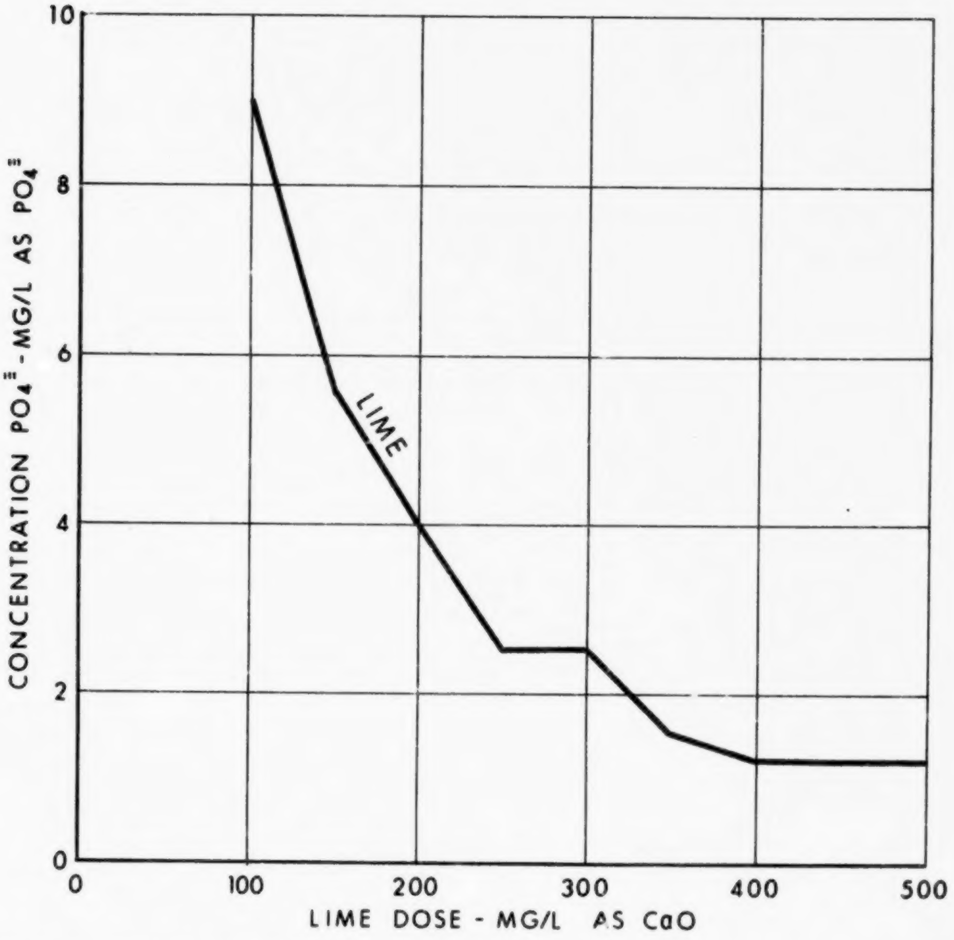
Lime Recalcination

The feasibility of recovering and reusing lime was evaluated in the laboratory by the schematic diagram shown on Figure 69. Four gallons of secondary effluent were dosed with 400 mg/l of lime (as CaO), flash-mixed for 1 minute and flocculated for 5 minutes. After 1 hour of settling, the clear liquor was decanted to another container, and the settled lime sludge was filtered through a Buchner funnel. The filtrate was added to the decanted liquor, and this mixture was recarbonated to a pH of 9.3. Four mg/l of polyelectrolyte was added to the recarbonated water to aid the settling of the fine dispersion of calcium carbonate floc. The resulting calcium carbonate sludge was filtered and the calcium content of the filtrate determined.

The lime and calcium carbonate sludges were recalcined in a muffle furnace at 1,800°F for 90 minutes. The calcium content of the recalcined sludges was determined, and the required amount of makeup lime was added to provide a dose of 400 mg/l as CaO for another four gallons of secondary effluent. This cycle was repeated eleven times.

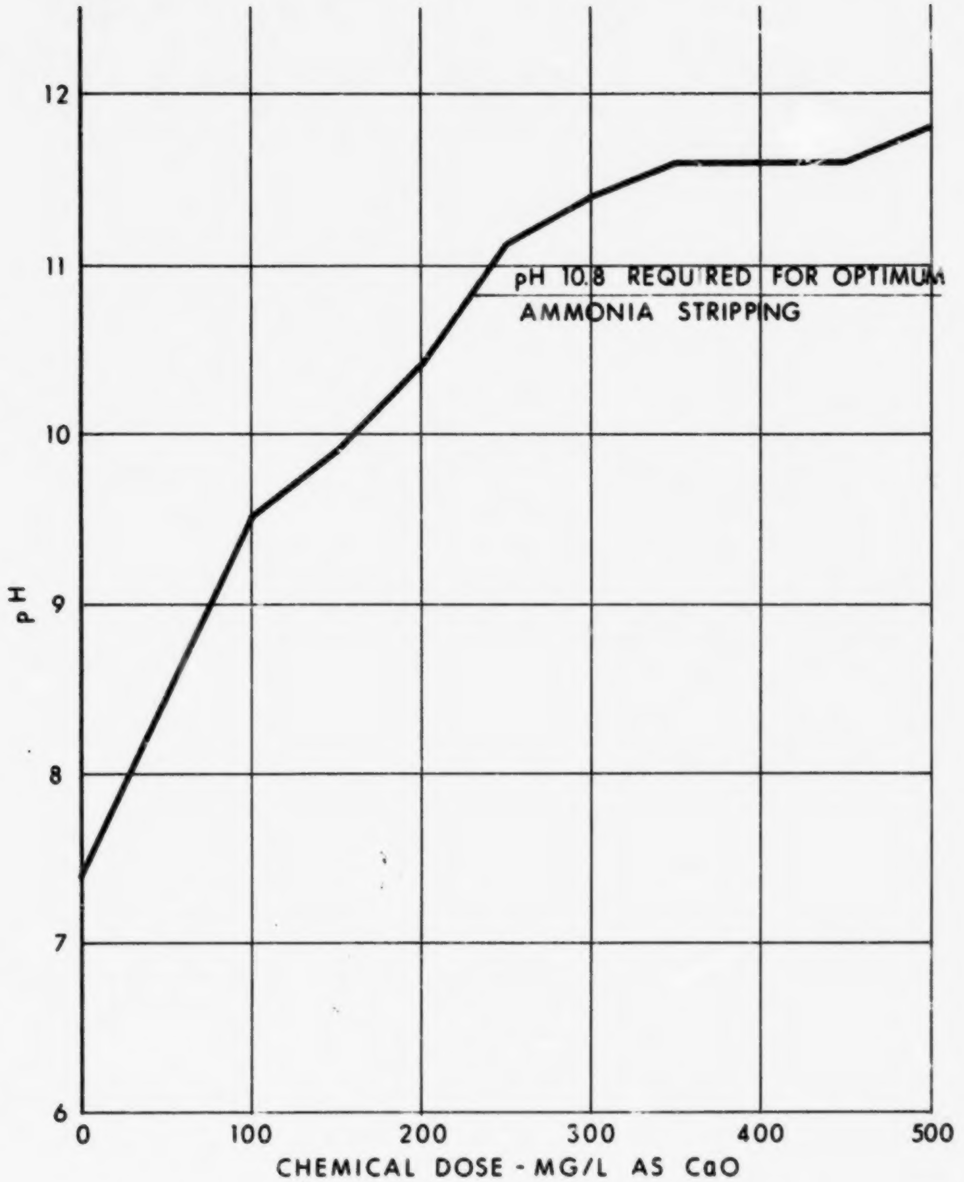
The percent makeup lime required to maintain a lime dosage of 400 mg/l as CaO showed little variation and averaged 36 percent over eleven cycles (Figure 70). However, the buildup of inert materials, such as calcium phosphate, in the recalcined ash resulted in a steady decrease in the percentage of CaO in the recalcined ash (Figure 71). This will necessitate wasting a portion of the reclaimed ash from the system to prevent the amount of material to be incinerated from increasing indefinitely and overtaxing the recalcining facilities.

Phosphate removal (Figure 72) was not adversely affected when the reclaimed lime was recycled. The average phosphate removal over eleven cycles was 97.5 percent, and the final phosphate concentrations, following settling, were less than 0.6 mg/l (as PO_4). In previous jar tests, Figure 69, the final phosphate concentration could only be reduced to 1.2 mg/l, with a lime dose of 400 mg/l. This slight increase in phosphate removal can be attributed to the use of larger samples having containers with better settling characteristics. Figure 73 shows that the recycled lime will provide a pH more than adequate for ammonia stripping.



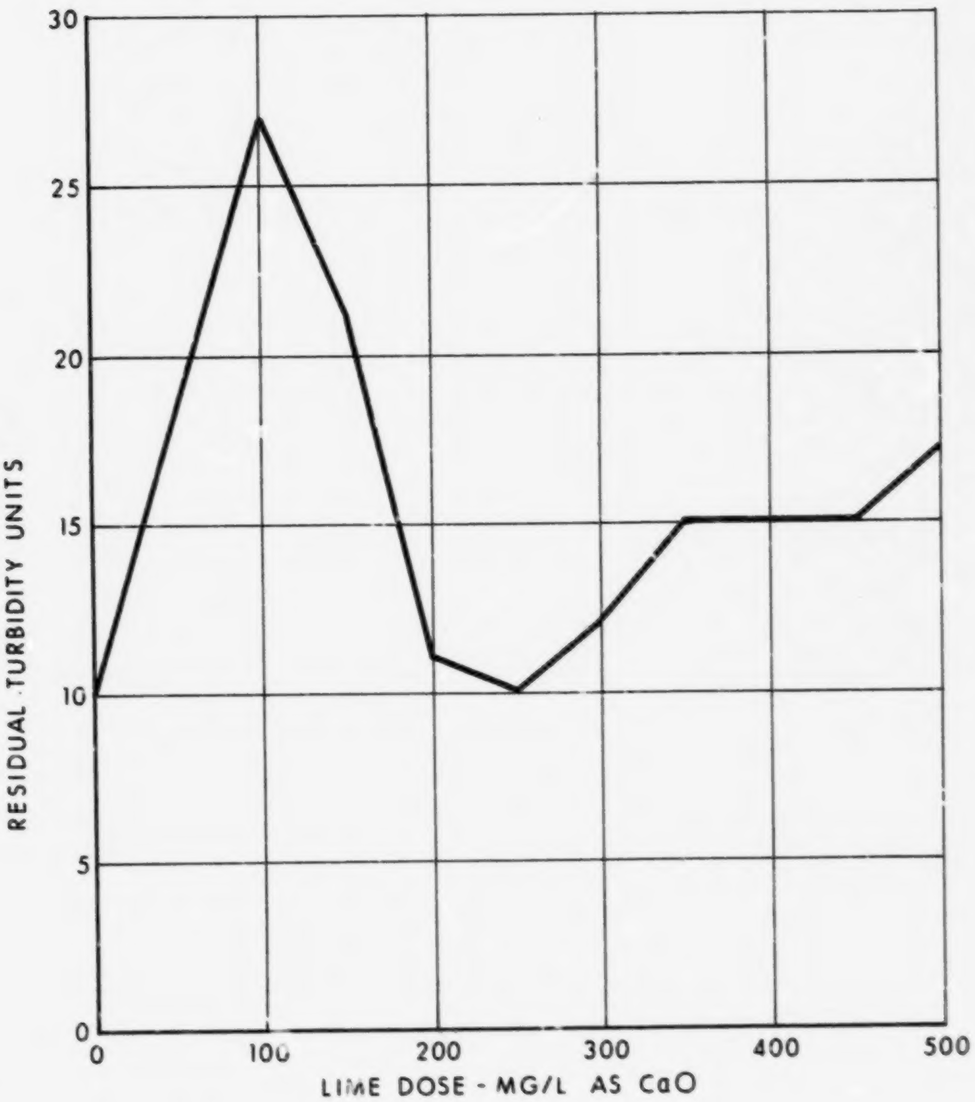
INITIAL $PO_4^{\equiv}$ = 21.5 MG/L

FIGURE 66
PHOSPHATE CONCENTRATION VS. LIME DOSE



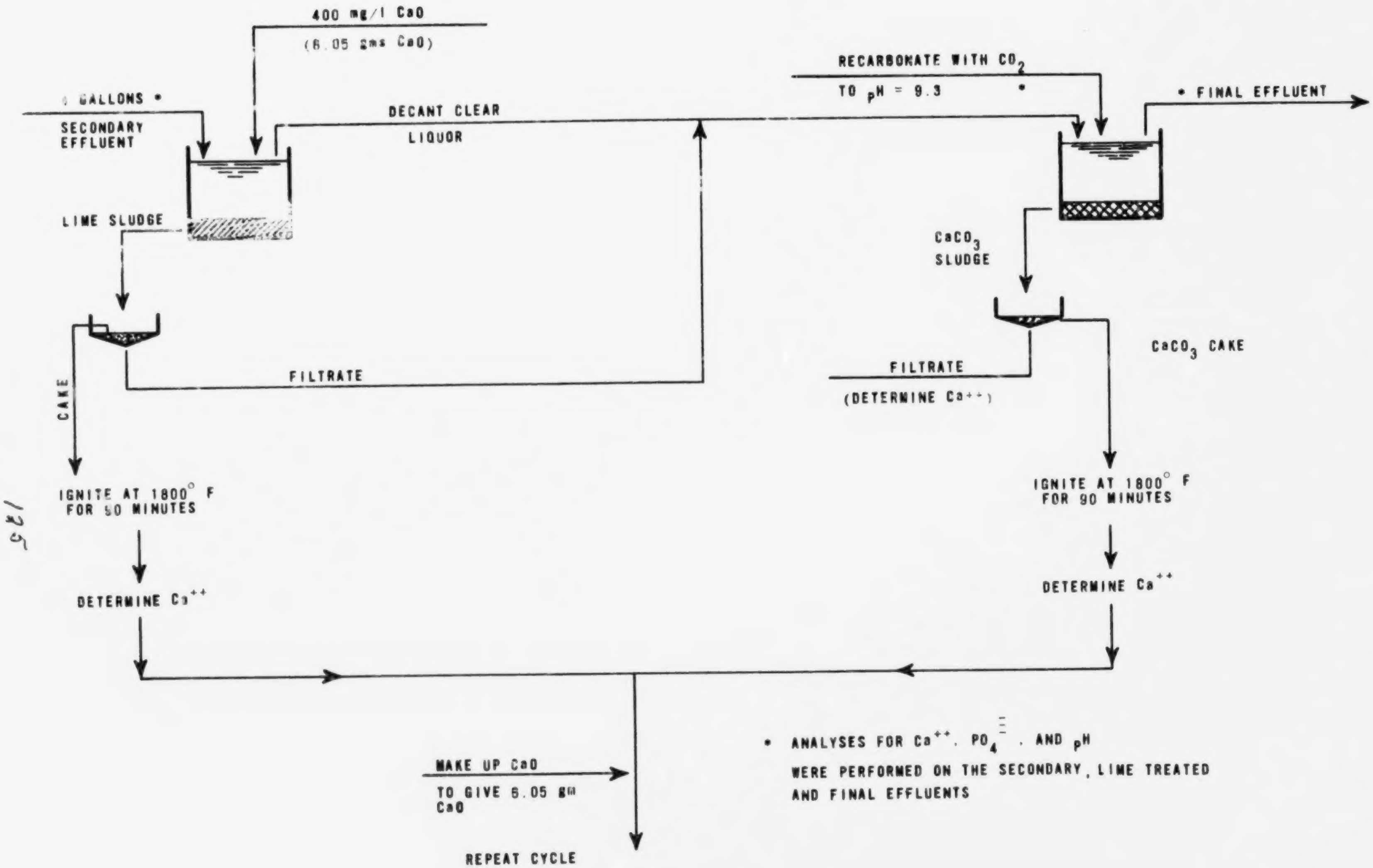
INITIAL pH = 7.4

FIGURE 67
pH VS. LIME DOSE



INITIAL TURBIDITY = 10

FIGURE 68
RESIDUAL TURBIDITY VS. LIME DOSE
(SETTLED BUT NOT FILTERED)


4 GALLONS •
SECONDARY EFFLUENT
400 mg/l CaO
(6.05 gms CaO)
DECANT CLEAR LIQUOR
RECARBONATE WITH CO_2
TO $_pH$ = 9.3 •
• FINAL EFFLUENT
LIME SLUDGE
FILTRATE
CAKE
$CaCO_3$ SLUDGE
FILTRATE
(DETERMINE Ca^{++})
$CaCO_3$ CAKE
IGNITE AT 1800° F FOR 50 MINUTES
DETERMINE Ca^{++}
IGNITE AT 1800° F FOR 90 MINUTES
DETERMINE Ca^{++}
MAKE UP CaO
TO GIVE 6.05 gm CaO
REPEAT CYCLE

• ANALYSES FOR Ca^{++}, $PO_4^{\equiv}$, AND $_pH$ WERE PERFORMED ON THE SECONDARY, LIME TREATED AND FINAL EFFLUENTS

FIGURE 69

SCHEMATIC OF LABORATORY PROCESS FOR EVALUATING LIME RECOVERY AND REUSE

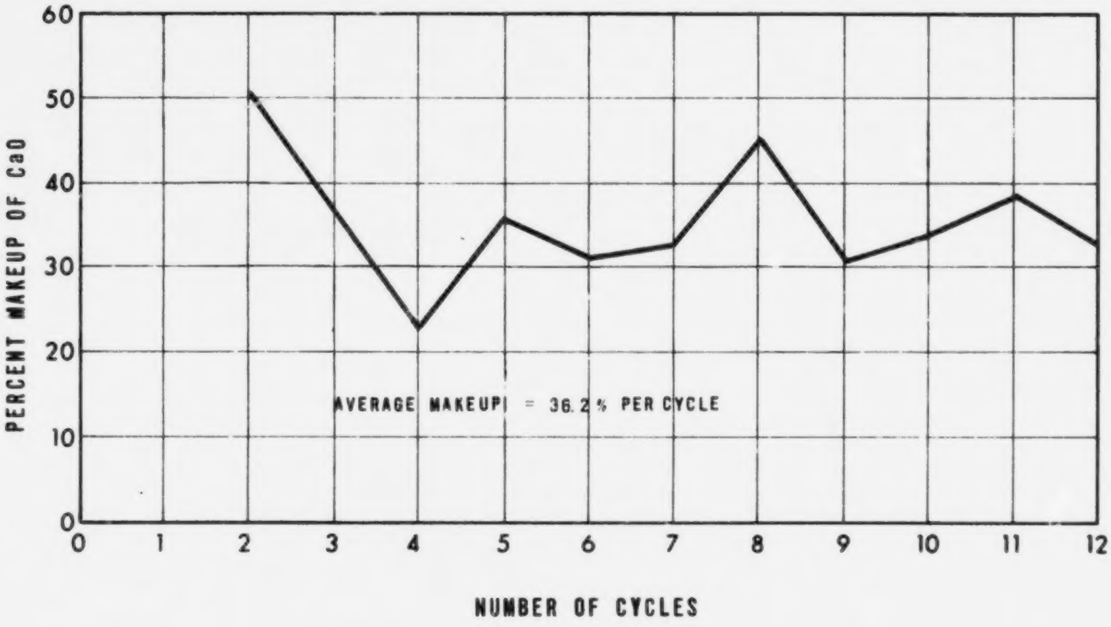


FIGURE 73

EFFECT OF RECYCLING ON PERCENT LIME MAKEUP

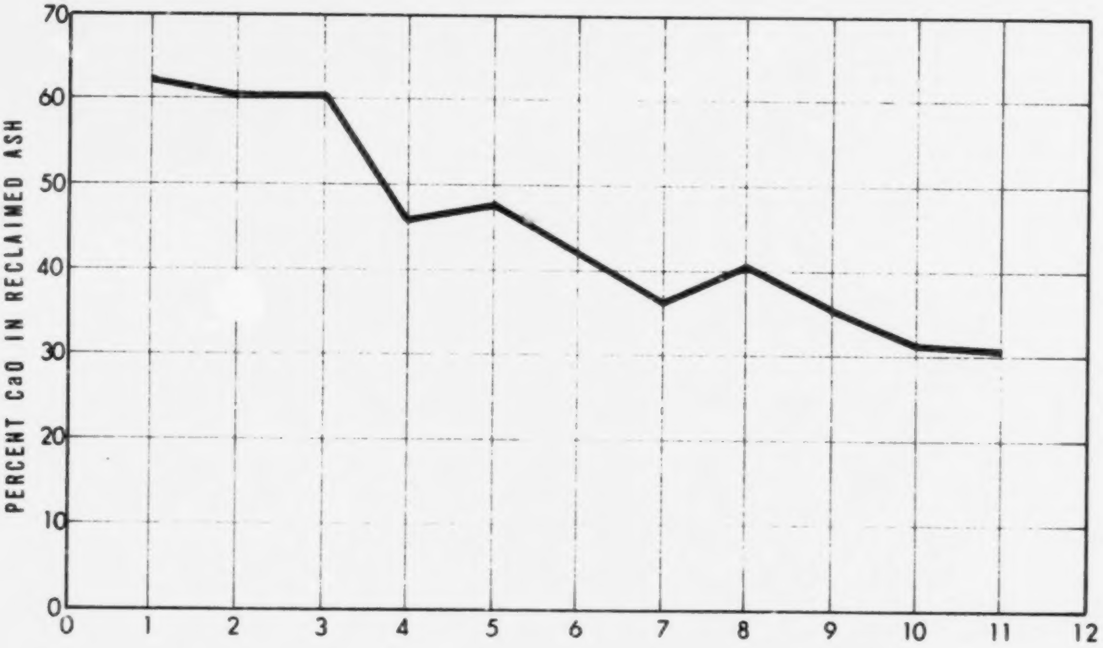


FIGURE 71

EFFECT OF RECYCLING LIME ON PERCENT CaO IN RECLAIMED ASH

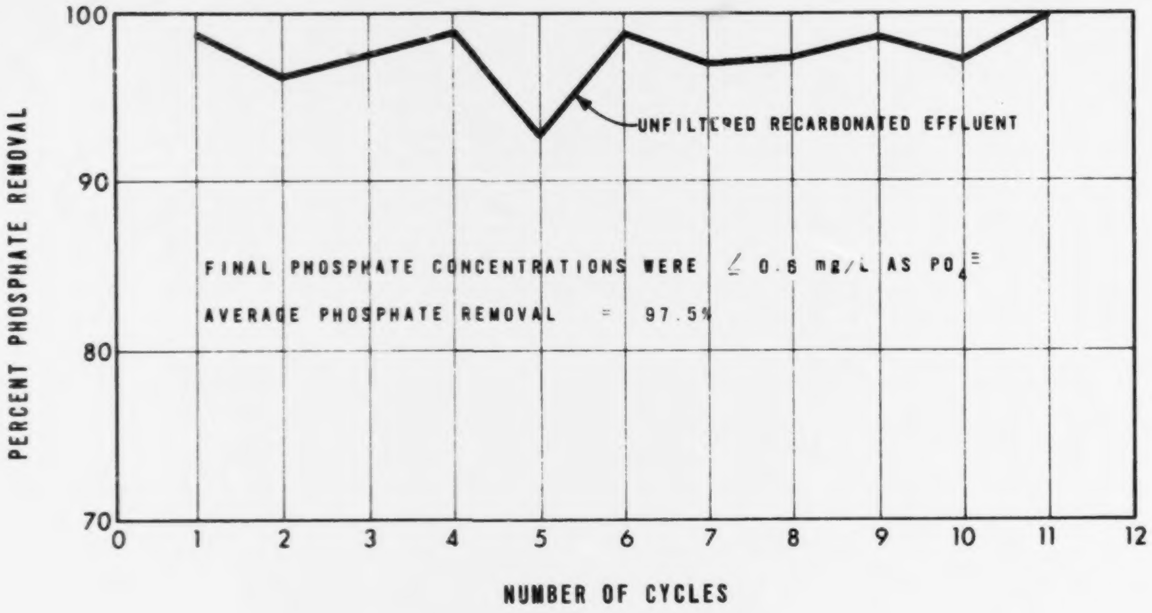


FIGURE 72

EFFECT OF RECYCLING LIME ON PHOSPHATE REMOVAL

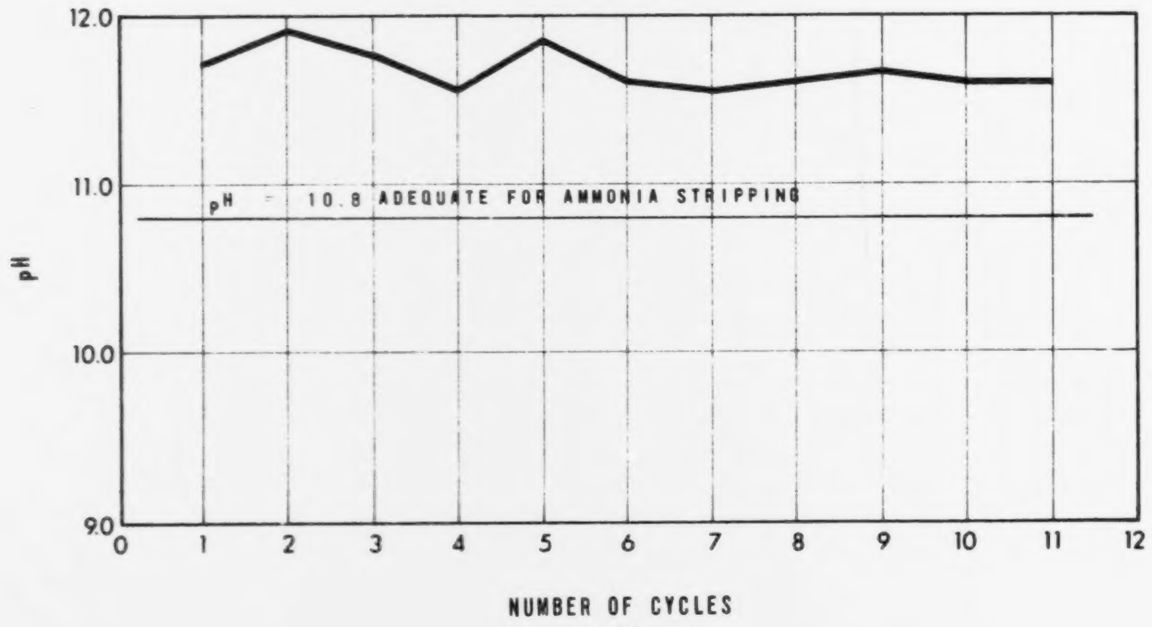


FIGURE 73

EFFECT OF RECYCLING LIME ON pH OF SEWAGE

There are several ways in which the wasting of reclaimed lime could be accomplished. The amount to be wasted could be hauled to a disposal area or added to the plant flow in the primary or secondary sedimentation basin. It was decided to waste the material to the primary sedimentation basin in an effort to remove some phosphate early in the process and raise the pH slightly, which will decrease the amount of makeup lime required.

Two studies were conducted in which 100 mg/l and 50 mg/l, respectively, were bled off to the primary sedimentation basin. The percent of makeup of CaO for each run is shown on Figure 74, with the data from the original study. The average makeup CaO was 48.0 percent and 42.9 percent, respectively, for 100 mg/l and 50 mg/l added to the primary.

As increasing amounts of reclaimed lime were bled off to the primary, the percent of CaO in the lime increased (Figure 75). It appears that a bleed-off of 100 mg/l will be sufficient to maintain the amount of reclaimed lime within the furnace capacity.

Figure 76 shows that phosphate removal was greater than 92 percent in all of the studies. Little difference from the initial study was noted when reclaimed lime was added to the primary.

The pH of the effluent increased slightly when increased amounts of reclaimed lime were added to the primary (Figure 77).

The effectiveness of pilot separation beds, identical to the mixed-media beds in the existing tertiary plant, in removing $CaCO_3$ floc from solution was evaluated, using lime-treated secondary effluent, recarbonated to pH 9.3. When the pilot separation beds were operated at a surface loading rate of 5 gpm/sq ft, the rate of total head loss increase varied from 1 to 2 feet per hour. Using a dose of 1 mg/l polyelectrolyte as a filtration aid, the separation bed effluent had a turbidity of 1 unit. There was essentially no pH drop across the separation beds, and no further precipitation of soluble calcium ion occurred in the beds.

When the chemical dosage applied to the separation beds was changed to 20 mg/l alum and 0.5 mg/l polyelectrolyte, turbidities of less than 0.5 units and phosphate concentrations less than 0.2 mg/l as PO_4 were obtained. The pH was reduced from 9.3 to 8.3 to 8.6. The separation bed influent calcium concentration was 50 mg/l (as Ca), and the effluent calcium concentration varied from 18 to 25 mg/l (as Ca), when using alum and polyelectrolyte. The calcium concentration of the secondary effluent before lime treatment was 15 mg/l as Ca. This indicates that the loss of calcium in the plant effluent would be less than 5 percent of the calcium added in coagulating secondary effluent.

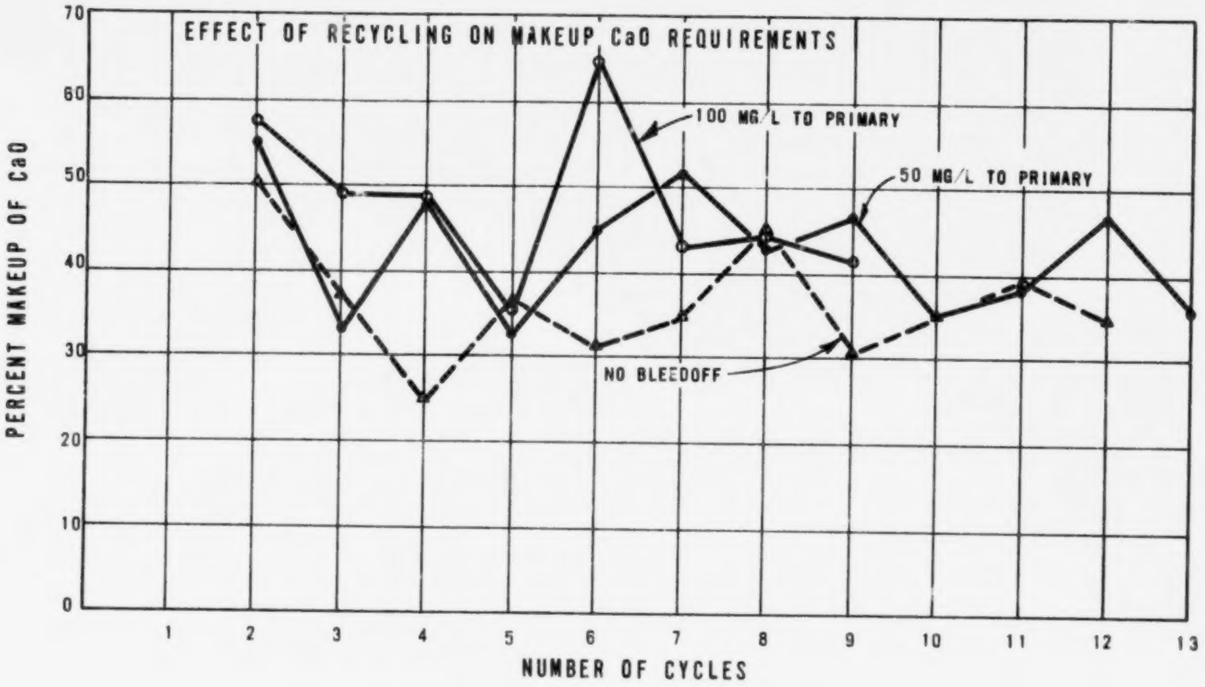


FIGURE 74

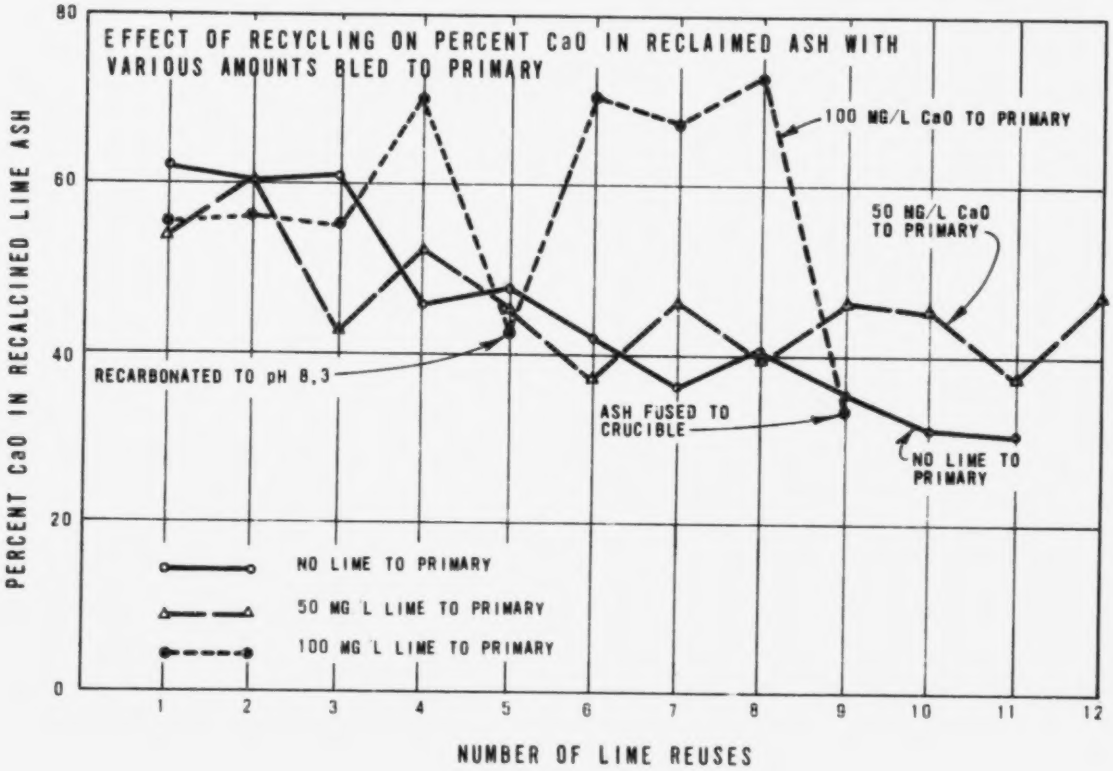


FIGURE 75

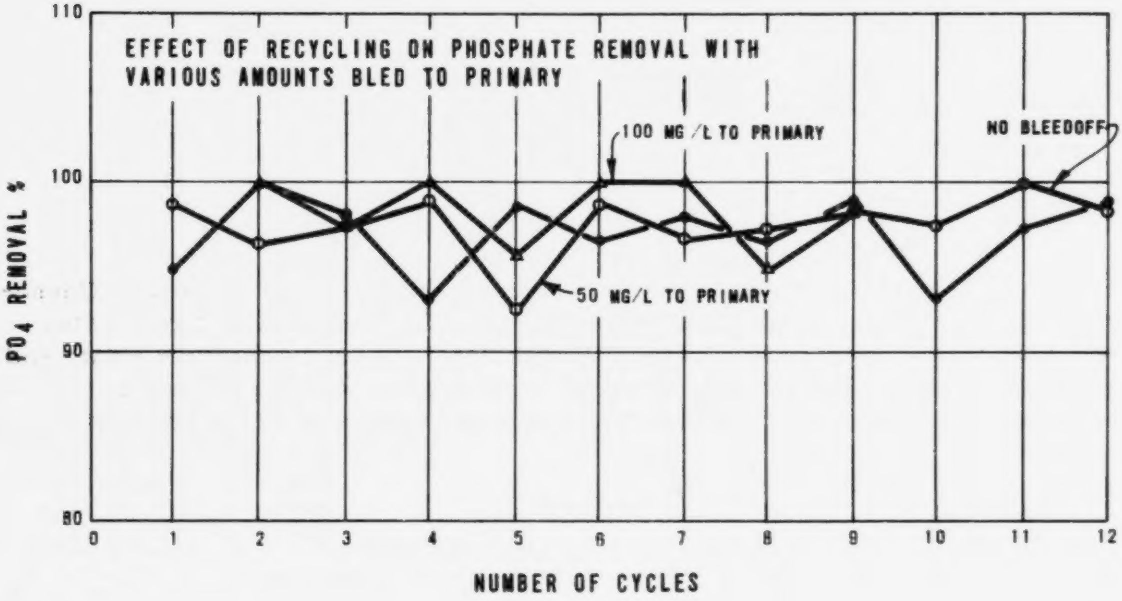
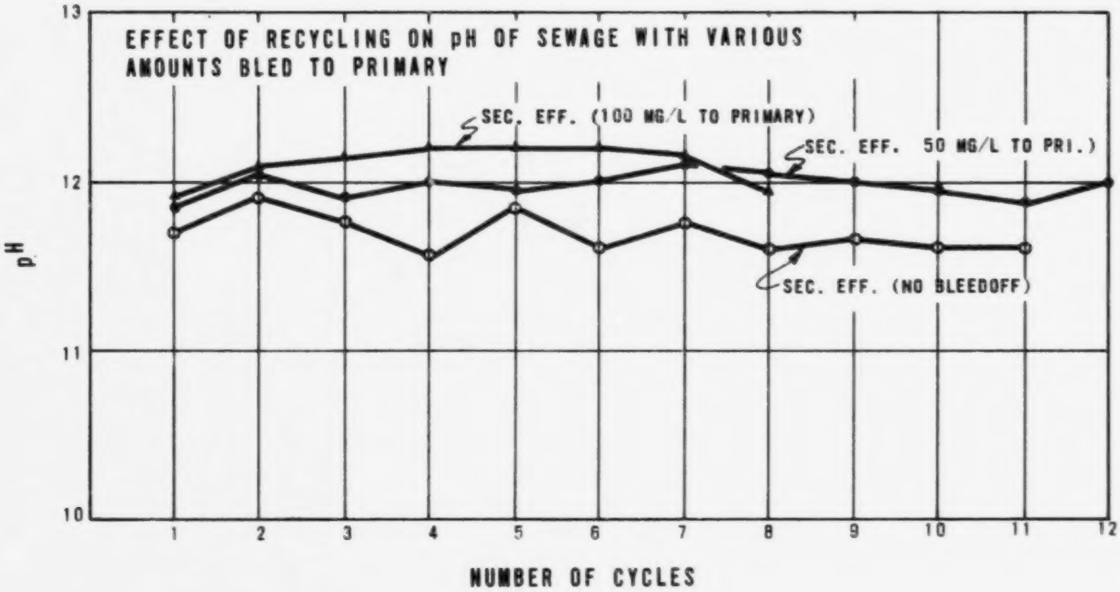


FIGURE 76



FIBURE 77

The calcium lost in the plant effluent, during the ash blown down and in the plant-scale recalcining operation, are estimated to bring the total makeup requirement for plant-scale treatment to 25 percent per cycle.

Cost of Lime Recovery

Table 18 presents a cost estimate for lime recovery at South Tahoe. The estimated total cost is shown to be $25.40 per ton, which is less than the $26.00 per ton cost of virgin lime delivered to South Tahoe. More importantly, the disposal problem created by large volumes of chemical sludge is eliminated.

Capital costs were amortized at 4 percent interest over 20 years. Chemical costs are f.o.b., South Lake Tahoe and include 4 percent California sales tax. Power was based upon a demand charge of $2.30 per kw hour per month and an energy charge of $0.008 per kw hour. Maintenance was based upon an annual cost of 3 percent of the equipment and labor costs of $3.75 per hour.

Table 18

ESTIMATED COST OF LIME RECOVERY
AT THE SOUTH TAHOE PUBLIC UTILITY DISTRICT
WATER RECLAMATION PLANT

	Cost Dollars/Ton Reclaimed CaO
Capital	$ 6.30
Fuel	7.35
Power	1.90
Maintenance	2.50
Labor	7.35
Estimated Cost of Lime Recovery/Ton Reclaimed CaO	$25.40

CHAPTER VII

METHODS FOR NITROGEN REMOVAL

GENERAL

The discharge of nitrogen-containing waste waters is considered by many to be an undesirable biological stimulant. Nitrogen with sufficient quantities of phosphorus may stimulate objectionable growths of algae and other aquatic plants. The tertiary treatment plant at the South Tahoe P. U. D., while removing over 98 percent of the phosphorus present in the secondary effluent, presently does not remove a significant amount of the nitrogen present.

Studies of various methods of nitrogen removal were performed at the District's treatment plant. These studies included an evaluation of the practicality, economics, and problems of plant-scale operation associated with the various methods. The methods of nitrogen removal studied included:

1. Microbial Denitrification
2. Ion Exchange
3. Ammonia Stripping
4. Chloramine Adsorption and Breakpoint Chlorination

MICROBIAL DENITRIFICATION

Theory

About 90 percent of the nitrogen in raw sewage is in the form of ammonia or urea from which ammonia is readily derived.(29) The other 10 percent is in the form of more complex organic materials, such as proteins and amino acids. Most of these more complex compounds break down to form ammonia. Provided oxygen is available, this ammonia may be converted to nitrates by bacteria found in the secondary sewage treatment processes. These nitrifying organisms (genera Nitrosomonas, Nitrosococcus, and Nitrobacter) exist autotrophically, obtaining their energy through the oxidation of inorganic substrate (ammonium ion and nitrite) to nitrate.

Provided aerobic conditions are maintained following nitrification, the nitrogen will remain in the form of nitrate. However, in the absence of oxygen, a large number of saprophytic bacteria, which constitute the bulk of the activated sludge organisms, can utilize nitrates as electron acceptors, making it possible for the organisms to grow in the absence of oxygen. Under these conditions, nitrates and nitrites are reduced to nitrogen gas, which effects a reduction in the nitrogen content of the waste water as it escapes from solution.

Literature Review

Essentially, all of the nitrogen present in the incoming waste water must be converted to nitrates if microbial denitrification is to provide efficient removal of nitrogen. For the activated sludge process, Johnson and Schroepfer(30) found the ratio of pounds of BOD added per day to pounds of volatile suspended solids in the aeration tank (load factor) to be the parameter which best described conditions for nitrification. Reasonably complete nitrification was obtained for load factors of 0.25 to 0.35. They considered it unlikely that a highly nitrified effluent would be obtained when the load factor was greater than 0.35. They found long aeration periods were not necessary for nitrification. A high degree of nitrification was obtained with aeration times as low as 2 to 3 hours when the load factor was less than 0.35. Wuhrmann(31) obtained a high degree of nitrification with aeration times of 1.7 to 2.2 hours when the mixed liquor suspended solids (MLSS) concentration (5,350 mg/l) was maintained at a level sufficient to obtain load factors less than 0.33. Wuhrmann(32) also found that high mixed liquor dissolved oxygen (D.O.) levels were not required to obtain nitrification. Essentially, complete nitrification was obtained with a mixed liquor D.O. of 1 mg/l when the load factor was 0.32.

Wuhrmann(32) found that the nitrates in a completely nitrified effluent from a municipal activated sludge plant could be greatly reduced by installing an anaerobic stage of treatment between the aeration basin and secondary clarifier. The amount of sludge present in the mixed liquor was adequate to provide denitrification provided there was a sufficient period of anaerobiosis. He found(31) that the mobilizable reserves of respiration substrates within the cells were large enough to permit the reduction of considerable quantities of nitrate without the addition of any supplemental substrate. Following 1.7 to 2.2 hours of aeration (MLSS = 5,350 mg/l) with a load factor of 0.25 to 0.33, a 2.2- to 2.8-hour period of anaerobiosis reduced the total nitrogen content from 20—25 mg/l to 3—4 mg/l.

Wuhrmann(31) has clearly demonstrated that no significant change of respiration rate (endogenous or substrate respiration) occurs when the activated sludge suffers anaerobiosis for several hours. Westgarth, et al(33) have also shown that prolonged periods of anaerobiosis (up to 10 hours) do not have a detrimental effect on the biological activity of activated sludge or on the overall efficiency of the conventional activated sludge process.

When trying to apply Wuhrmann's process in a laboratory study using a synthetic sewage feed, Johnson and Schroepfer(30) found the reduction of nitrate without addition of raw waste to the denitrification unit to be impracticable because of the slow rate of denitrification. Wuhrmann(32) did not find it necessary to add raw sewage in his studies. Raw sewage made up 20—40 percent of the flow to the denitrification unit in the study of Johnson and Schroepfer. They found that 20 percent raw waste gave complete denitrification in about 3 hours. Greater proportions of raw waste did not appear to increase the rate of denitrification. They also found the MLSS concentration did not appear to have any noticeable effect on the rate of denitrification. These observations are difficult to explain because the oxygen demand in the denitrification unit should be proportional to the concentration of organisms present and the amount of raw waste added.

Although the addition of raw waste to the denitrification unit increased the rate of denitrification in the studies of Johnson and Schroepfer, it also decreased the efficiency of total nitrogen removal. This is to be expected because a substantial amount of unoxidized nitrogen is added to the denitrification unit when the raw waste is added. The addition of raw waste reduced the efficiency from the 90 percent found by Wuhrmann to 54—65 percent. A nitrogen balance(30) showed that essentially all of the incoming nitrate was reduced to nitrogen gas.

Ludzack and Ettinger(34) developed a "semi-aerobic" process in the laboratory in an attempt to minimize activated sludge effluent nitrogen. The aeration in the influent zone of the aeration unit was greatly reduced. Nitrified mixed liquor from later stages of aeration was added to this zone to provide a source of nitrate oxygen. Aeration in this zone was held to a level just sufficient for solids suspension. The fact that the D.O. reached values of up to 0.5 mg/l in this denitrification zone undoubtedly reduced the rate of denitrification. They found variations in flow rate caused difficulty in maintaining the D.O. low enough for effective denitrification in the influent zone. Although it has been shown(31,33) that long periods of anaerobiosis do not adversely affect activated sludge, they were concerned that the sludge not remain septic for long periods. As would be expected under these less-than-optimum environmental conditions for denitrification, the efficiency of the "semi-aerobic" process was considerably less, 20—77 percent total nitrogen removal, than that achieved by Wuhrmann's process.

Eliassen and Wyckoff(35) and Barth, et al(36) also report that conditions are less than optimum for denitrification when the denitrification chamber is part of the mixed liquor compartment. After constructing a separate pilot-scale denitrification unit, Eliassen and Wyckoff(35) found optimum nitrogen removal was obtained when the load ratio on the aeration tank was 0.21 to 0.29, and the detention period in the denitrification tank was 47 to 78 minutes. During this period of optimum operation, 76 to 77 percent nitrogen reduction was obtained. A wide fluctuation in efficiency occurred during the remainder of their study, with nitrogen removals varying from 27 to 84 percent.

Although Wuhrmann(32) feels that microbial denitrification would be more widely applicable than other proposed methods of nitrogen removal, he points out(31) that the resolution of the problem of nitrogen removal by denitrification remains economically difficult since complete nitrification of the incoming waste water may be expensive. Also, it is clear that the efficiency of the process is subject to wide fluctuations. McKinney(37) feels that the denitrification process will not find extensive use in practice due to these fluctuations in efficiency and the cost of the process.

Experimental Procedures

A nitrified effluent was obtained by aerating primary effluent from the South Tahoe P. U. D. Water Reclamation Plant in a 145-gallon pilot plant aeration tank. The load factor of the full-scale, activated sludge plant aeration tank varied from 0.3 to 0.4 during the study period. No significant degree of nitrification occurred in the full-scale plant under these conditions.

The detention time and MLSS level of the pilot aeration tank were controlled to maintain a load factor of 0.1 to 0.3. Nitrified effluent was not obtained when the load factor greatly exceeded 0.3. This is consistent with the observations in the full-scale plant.

In order to evaluate the parameters which would affect the design of a denitrification unit, laboratory tests were conducted with 500-ml portions of the nitrified mixed liquor. A laboratory gang mixer was used to mix these samples of mixed liquor during the period of anaerobiosis required for denitrification. The MLSS and amount of raw waste added to the nitrified mixed liquor were varied.

Experimental Results

Contrary to the findings of Johnson and Schroepfer,(30) the MLSS concentration was found to affect the rate of denitrification. As shown on Figure 78, the rate of denitrification increased significantly as the MLSS was increased from 1,130 mg/l to 4,770 mg/l. The rate of oxygen demand increases as the concentration of organisms increases; thus, the results shown on Figure 78 are not surprising.

The effects of the addition of raw sewage upon the rate of denitrification for two different MLSS levels are shown on Figure 79. The rate of denitrification was increased significantly by the addition of raw sewage. The rate increased as the proportion of raw sewage was increased from 20 percent to 40 percent. Although the rate of denitrification increased greatly with the addition of raw sewage, the minimum effluent nitrogen content was obtained with high MLSS (4,470 mg/l) and no addition of raw sewage, as shown in Table 19.

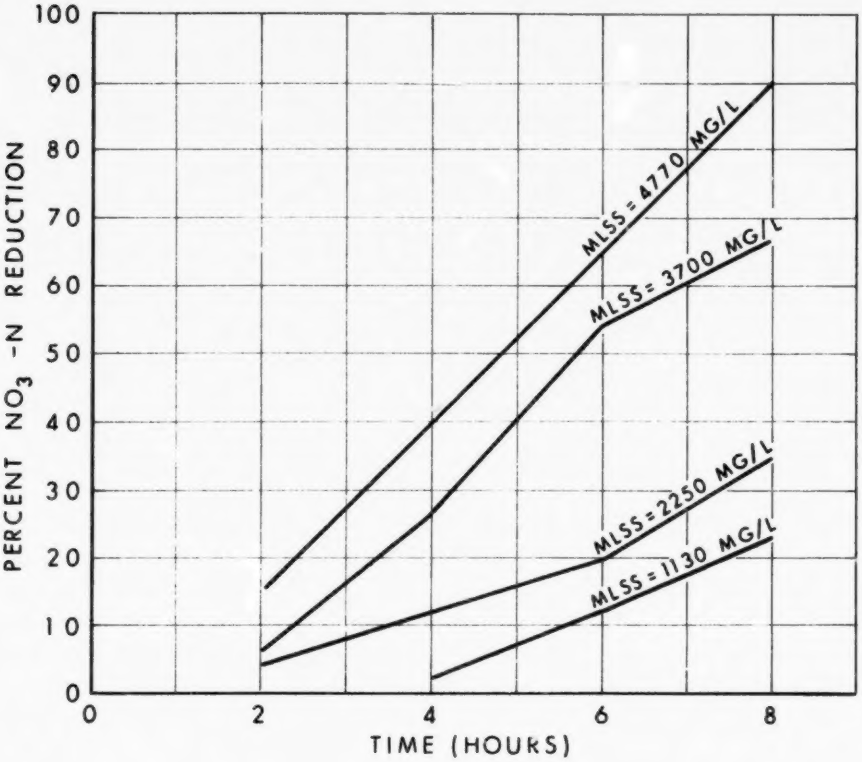


FIGURE 78
EFFECT OF MIXED LIQUOR SUSPENDED SOLIDS CONCENTRATION ON RATE OF DENITRIFICATION (NO RAW SEWAGE ADDED)

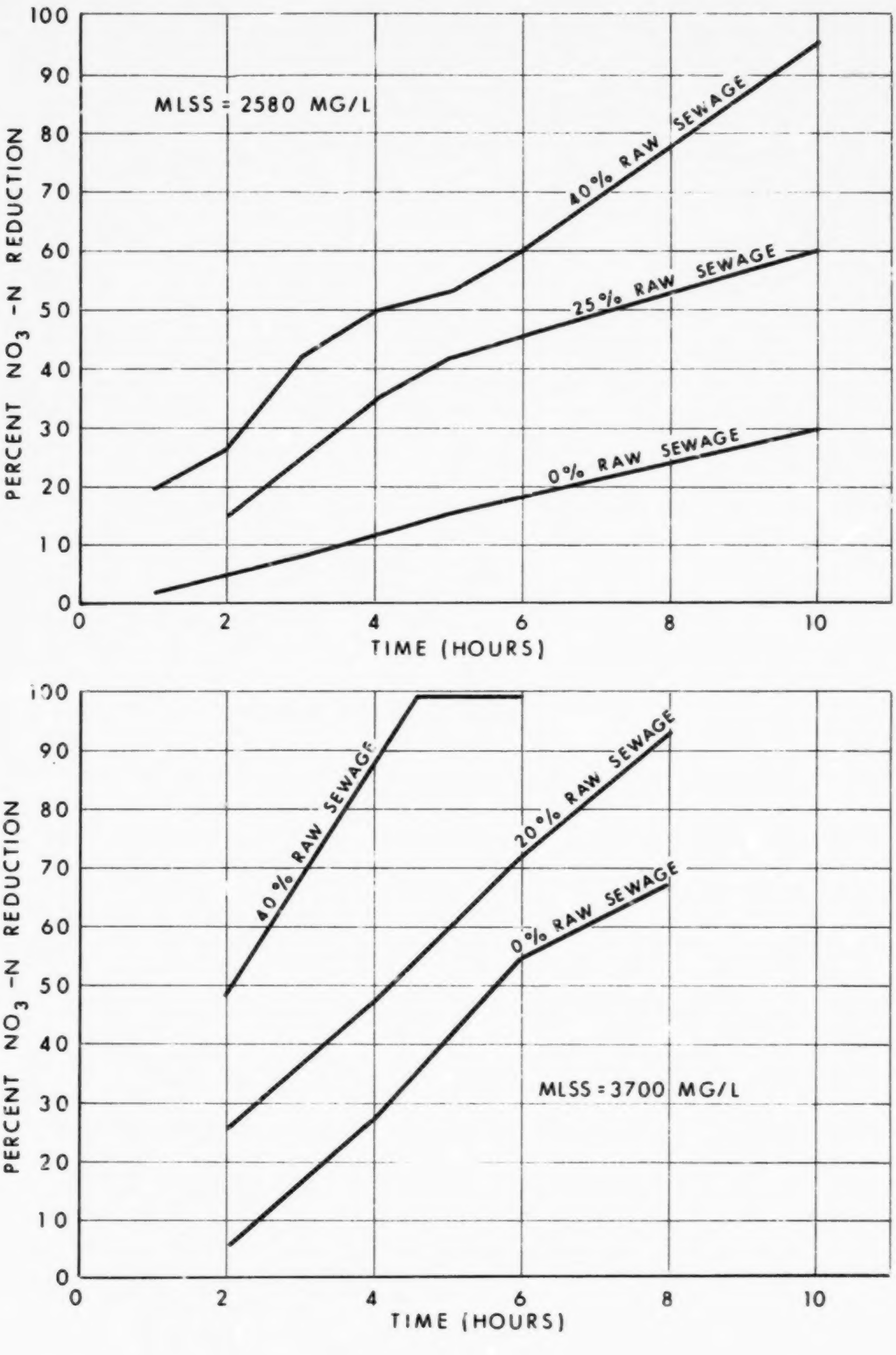


FIGURE 79
EFFECT OF RAW SEWAGE ADDITION ON RATE
OF DENITRIFICATION

Table 19

EFFICIENCY OF NITROGEN REMOVAL BY THE DENITRIFICATION PROCESS

MLSS	Detention Time (hrs)	Percent Raw Sewage	Total Nitrogen (mg/l) Influent	Total Nitrogen (mg/l) Effluent	Percent Nitrogen Removal
1130	8	0	55	43	22
2250	8	0	55	38	31
4470	8	0	55	10	82
2580	10	0	54	42	22
2580	10	25	54	29	28
2580	10	40	50	18	64
3700	8	0	59	29	51
3700	8	20	55	20	64
3700	6	40	49	20	59

The sewage entering the aeration unit contained about 50–55 mg/l total nitrogen, approximately twice that found in most municipal waste waters. Ammonia-nitrogen comprised 30–40 mg/l of this total, with essentially all of the remainder of the incoming nitrogen in the form of organic nitrogen. If 40 percent of this sewage were placed in the denitrification unit, approximately 20 mg/l of unoxidized nitrogen would be added and could be expected to appear in the effluent from the denitrification unit, even with complete denitrification. This is illustrated by the data obtained from the addition of 40 percent raw sewage to the nitrified mixed liquor, with a MLSS of 3,700 mg/l (see Table 19 and Figure 79). Complete denitrification was obtained in slightly over 4 hours, but the percent total nitrogen removal was only 59 percent, because of the residual ammonia and organic nitrogen remaining from the addition of raw waste.

As this study, previous studies,(30,31,32) and logic would indicate, maximum nitrogen removals will be obtained when no raw waste is added to the denitrification unit while the maximum rate of denitrification will be obtained with the addition of raw sewage. If nitrogen removals of about 60 percent are adequate, the size of the aeration and denitrification units could be significantly reduced by bypassing a portion of the raw sewage around the aeration unit and placing it in the denitrification unit. However, when nitrogen removals on the order of 80–95 percent are required, it would be necessary to depend solely on the oxygen demand exerted by the MLSS for denitrification.

The tertiary treatment process in use at the South Tahoe P. U. D. plant consists of coagulation of the secondary effluent with alum and filtration through multimedia beds, followed by filtration through granular-activated carbon columns. If the denitrification process were to be applied ahead of this tertiary process, the addition of raw waste to the denitrification unit would result in more soluble organic matter reaching the activated carbon columns. This would, of course, increase the cost of tertiary treatment. Thus, while the addition of raw waste might provide significant savings at some plants, the savings provided at the South Tahoe plant would be offset to some degree by an increase in the cost of tertiary treatment. The cost estimates developed for the denitrification process and the other methods of nitrogen removal being studied are based upon total nitrogen removals of 80—95 percent.

The laboratory studies indicate that, with a completely nitrified mixed liquor, a period of anaerobiosis of about 8 hours with a MLSS concentration of about 5,000 mg/l would result in a total nitrogen removal of 85—90 percent at the South Tahoe P. U. D. plant. The oxygen demand of the mixed liquor solids at the end of the aeration period was found to be 4.9 mg oxygen per gram of dry solids per hour. On this basis, with 5,000 mg/l MLSS, it would theoretically require 9 hours to remove 50 mg/l nitrate nitrogen with the oxygen demand exerted by the mixed liquor solids. Thus, the laboratory observations correlate well with the theoretical rate of denitrification.

The detention time found to be required for denitrification is significantly longer than the 2.2 to 2.8 hours reported by Wuhrmann. This may be explained by the fact that the oxygen demand of the mixed liquor measured during these studies is about 30 percent of that observed by Wuhrmann(32) during his studies of denitrification. The higher oxygen demand of the mixed liquor solids he observed, coupled with a nitrate concentration of only about 50 percent of that experienced at the South Tahoe P. U. D. plant, account for the lower detention time Wuhrmann found satisfactory for denitrification with MLSS of slightly more than 5,000 mg/l.

Feasibility of Process

As previously discussed, a basin providing approximately 8 hours detention would be required for denitrification with a MLSS of 5,000 mg/l. For the purposes of this cost estimate, a plant capacity of 5 mgd has been assumed. A denitrification basin of 1.65-million-gallon capacity, equipped with mechanical mixers, would cost an estimated $313,600 to construct at Tahoe (including engineering fees and contingencies).

The volatile fraction of the MLSS at the South Tahoe P. U. D. plant has averaged 72 percent. Thus, to maintain a load factor of 0.25, with an assumed MLSS of 5,000 mg/l and a primary effluent BOD of 200 mg/l, a total aeration tank volume of 1.1 million gallons would be required. The existing aeration tanks have a volume of 0.9 million gallons; thus, an additional aeration basin with a volume of 0.2 million gallons would be required. It would cost an estimated $75,000 to construct the additional aeration basin at South Tahoe.

Additional sludge pumping facilities and air compressors also would be required in order to produce a nitrified effluent. The total estimated cost (above that required for conventional treatment) to include facilities for nitrification and denitrification in a 5-mgd plant at Tahoe is $454,000. If this capital cost were amortized over a 20-year period at 4 percent interest, the capital cost, based on a daily flow of 5 mgd, would be $18 per million gallons.

In order to maintain a MLSS of 5,000 mg/l, a higher rate of sludge return than used to maintain the normal MLSS at the South Tahoe P. U. D. plant would be required. The cost of the additional power required to return the greater sludge volumes, the power required for the denitrification mixers, and the additional air for oxidation of the nitrogen would cost an estimated $14 per million gallons. Table 20 summarizes the estimated capital and power cost of providing nitrification and denitrification (the cost above that required for conventional treatment) at the South Tahoe P. U. D. plant.

Table 20

ESTIMATED POWER AND CAPITAL COSTS
FOR NITRIFICATION AND DENITRIFICATION AT THE
SOUTH TAHOE P. U. D. WATER RECLAMATION PLANT

Item	Estimated Cost for 5-mgd Flow ($ per million gallons)
Capital cost of $454,000, amortized over 20 yrs at 4 percent interest	$18
Power	14
Total	$32

ION EXCHANGE

Theory

Ion exchange may be defined as the reversible interchange of ions between a liquid and a solid, involving no radical change in the structure of the solid.(38) Resinous exchangers are insoluble, high molecular weight compounds containing acidic or basic groups, depending upon whether their purpose is to remove cations or anions. Because the nitrogen to be removed by ion exchange in these studies was in the form of ammonia, the cationic resins were of prime concern.

Cation exchangers in the hydrogen form contain reactive acid groups as part of the resin molecule. Other cations, such as the commonly used sodium, can be used in place of the hydrogen ion in these cation exchangers. Regeneration with sodium chloride, an inexpensive regenerant, activates the cation exchanger in the sodium form. An equation representing the reaction of ammonia-nitrogen and a cationic resin in the sodium form is as follows:(38)

$$2\,NH_4^+ + Na_2R \underset{\text{regeneration}}{\overset{\text{reaction}}{\rightleftarrows}} (NH_4)_2R + 2\,Na^+$$

Because ion exchange is controlled by electrostatic forces, the relative adsorbability of various ions is dependent upon the relative charges and ionic radii of the ions. Thus, divalent ions will be preferentially adsorbed relative to the monovalent ammonium ion. In the dilute concentrations found in sewage, the following represents the order in which cation exchange will occur.(38)

$$Ca^{++} < Mg^{++} < NH_4^+ < K^+ < Na^+$$

The above relation makes it evident that a high hardness would reduce the effective exchange capacity of the resin for ammonium ion removal. Fortunately, the sewage being treated in these studies had a low total hardness, about 55 mg/l.

Literature Review

A standard use of the ion exchange process is the removal of inorganic ions from water, as evidenced by the large number of water softening and deionization plants throughout the world. In applying the process to a tertiary sewage effluent, the goal is to remove inorganic nutrients. Since nearly all the phosphorus is removed by the alum treatment employed at the South Tahoe P. U. D. plant, the prime concern in this study was the removal of nitrogen by ion exchange.

Many natural and synthetic materials exhibit ion exchange properties. Natural aluminosilicates, called zeolites, have been used commercially for many years as ion exchange materials.(39) In recent years, synthetic ion exchange materials have been used extensively. These synthetic exchangers are insoluble, three-dimensional networks, which have sites within the structure where an ion is released and another attached.(39)

Many organic materials also may be removed by ion exchange and may be difficult to remove from the resin. This organic fouling of the resin hindered many of the previous attempts to treat sewage effluents by ion exchange. The tertiary process used at Lake Tahoe removed about 85 percent of the organic material found in secondary effluent. Thus, the organic load on the ion exchange resin is greatly reduced.

Another factor which further reduces the organic fouling problem is the development of porous resins which have the ability to reversibly adsorb organic materials. In the early synthesis of ion exchangers, much emphasis was placed on obtaining the maximum number of exchange groups per unit of volume, and little attention was paid to the porosity of the resin.(40) As a result, many of the early synthetic resins had high initial capacities which declined rapidly, because of fouling by ions which could not diffuse freely out of the interior of the resin.

Previous experimental applications of ion exchange for sewage treatment have dealt primarily with treatment of secondary effluent. One exception is the study by Rand and Nemerow,(28) in which presettled raw sewage was passed through an anion exchange material (Resex 42). Excellent reductions of ABS, BOD, and phosphate were obtained, but physical plugging of the bed limited the length of run to 62 bed volumes. Only one run was reported, and no study of regeneration of the resin was made.

Perhaps the first reported attempt to remove nitrogen from sewage by ion exchange is that reported by Gleason and Loonam(41) in 1933. The process consisted of four major steps:

1. Removal of suspended solids from raw sewage by coagulation with lime and iron compounds.
2. Sludge disposal by vacuum filtration and incineration.
3. Removal of nitrogen compounds by ion exchange with zeolite.
4. Regeneration of the zeolite with a salt solution and the subsequent recovery of ammonia from this solution.

The nitrogen, in the form of ammonia, was exchanged for sodium. The zeolite was regenerated with a salt solution containing 20 percent sodium chloride. Salt solution requirements were equal to 1 percent of the throughput. Approximately 16,000 pounds of zeolite were required per million gallons of sewage treated. It was estimated that such a plant, with a 50-mgd capacity, could be installed at about two thirds the cost of an activated sludge plant. Unfortunately, no data were presented to show the initial exchange capacity of the resin and the exchange capacity after several regenerations.

Nesselson(38) studied the use of ion exchange on secondary sewage effluent which had been passed through an anthracite coal filter. Strong base anion exchangers were found to satisfactorily remove nitrate by chloride anion exchange, but the handling of the large volume of waste regenerant (about 7 percent of the throughput volume) presented a difficult problem. He found the anion exchange capacity at a low flow rate, 1.2 gpm/cu ft, to be only 18 percent less than that at a high flow rate, 25 gpm/cu ft. Approximately 150 bed volumes of effluent were treated, on the average, prior to the effluent NO_3-N concentration increasing to 1 mg/l. Regeneration was accomplished with 40 to 70 pounds of sodium chloride per cubic foot of anion exchange resin.

Nesselson(38) studied the use of nuclear sulfonic cation exchangers for the removal of ammonia-nitrogen. Only a small portion of the cations present in the sewage effluent were ammonium ions. Thus, a large portion of the exchange capacity of the resin was being used to remove calcium and magnesium. Pretreatment of the effluent with lime to reduce the calcium and magnesium content increased the ammonium exchange capacity by 30 percent. Approximately 100 bed volumes were passed through the resin, on the average, before the NH_3-N increased to 1 mg/l. Regeneration was accomplished with 20 pounds of salt per cubic foot of resin, and the waste regenerant volume averaged about 6 percent of the throughput volume. It was estimated that the cost of the salt required for regeneration would be $260 per million gallons of effluent treated.

Eliassen and Wyckoff(35) applied sand-filtered secondary effluent to anion exchange resin for nitrate and phosphate removal. Removals were good (75 percent nitrate removal and 97 percent phosphate removal) to over 200 bed volumes passed. Significant amounts of COD, ABS, and color were also removed. The efficiency of the resin was reduced by a high chloride and sulfate concentration in the sewage being treated. Organic fouling was also occurring, as evidenced by a decreasing throughput to phosphate breakthrough with continued use of the resin. A method for removing the organic material from the resin was being studied. Regeneration of the resin was accomplished with 2 bed volumes of 10 percent sodium chloride. It was estimated that the cost of salt for regeneration would be $40 per million gallons of effluent treated. An attempt was made to apply unfiltered secondary effluent directly to the resin, but the bed plugged rapidly, and it was difficult to completely clean the resin by backwashing.

Eliassen, Wyckoff, and Tonkin(42) attempted to apply their laboratory scale ion exchange system(35) on a larger pilot scale. Phosphate removal was only about 70 percent on the pilot scale, rather than the 97 percent obtained in the laboratory studies. However, nitrate removal was higher (90 percent) in the pilot-scale unit. The authors attributed the differences in the laboratory and pilot-scale units to a higher chloride concentration in the sewage and to mechanical problems encountered in these preliminary runs with the pilot plant.

Experimental Procedures

Because the nitrogen to be removed at the South Tahoe plant is in the form of ammonia, a cationic resin was selected for these studies. The resin used was Duolite C25, furnished by Diamond Alkali Company. Duolite C25 is a tan, opaque, bead-form polystyrene-sulfonic acid resin. The C25 resin is a very porous resin designed to minimize organic fouling.

A 4-inch-diameter plexiglass cylinder was filled to about mid-depth with 0.20 cubic foot of C25 resin. The resin depth was 27.5 inches. Secondary effluent which had received coagulation with alum, filtration, and adsorption on activated carbon served as the influent for the ion exchange resin. The quality of the ion exchange influent is summarized in Table 21. The total calcium and magnesium concentration is much lower than the 385 mg/l in the sewage subjected to cation exchange by Nesselson.(38)

The flow of carbon column effluent to the ion exchange column was regulated by a rotameter. A schematic diagram of the ion exchange pilot plant is shown on Figure 80.

Table 21

QUALITY OF ION EXCHANGE INFLUENT

Parameter	Concentration
Biochemical Oxygen Demand, mg/l	Less than 1
Suspended Solids, mg/l	Less than 0.5
Chemical Oxygen Demand, mg/l	3—25
Turbidity, mg/l	Less than 1
Orthophosphate, mg/l as PO_4	Less than 1
ABS, mg/l	Less than 0.5
Color, units	Less than 5
Alkalinity, mg/l as $CaCO_3$	70
Organic Nitrogen, mg/l as N	2
Ammonia-Nitrogen, mg/l as N	18—28
pH	6.4—6.8
Calcium, mg/l	16
Magnesium, mg/l	5
Sodium, mg/l	60
Chlorides, mg/l	35
Sulfates, mg/l	210
Total Hardness, mg/l	55
Specific Conductance, micromhos/cm	425

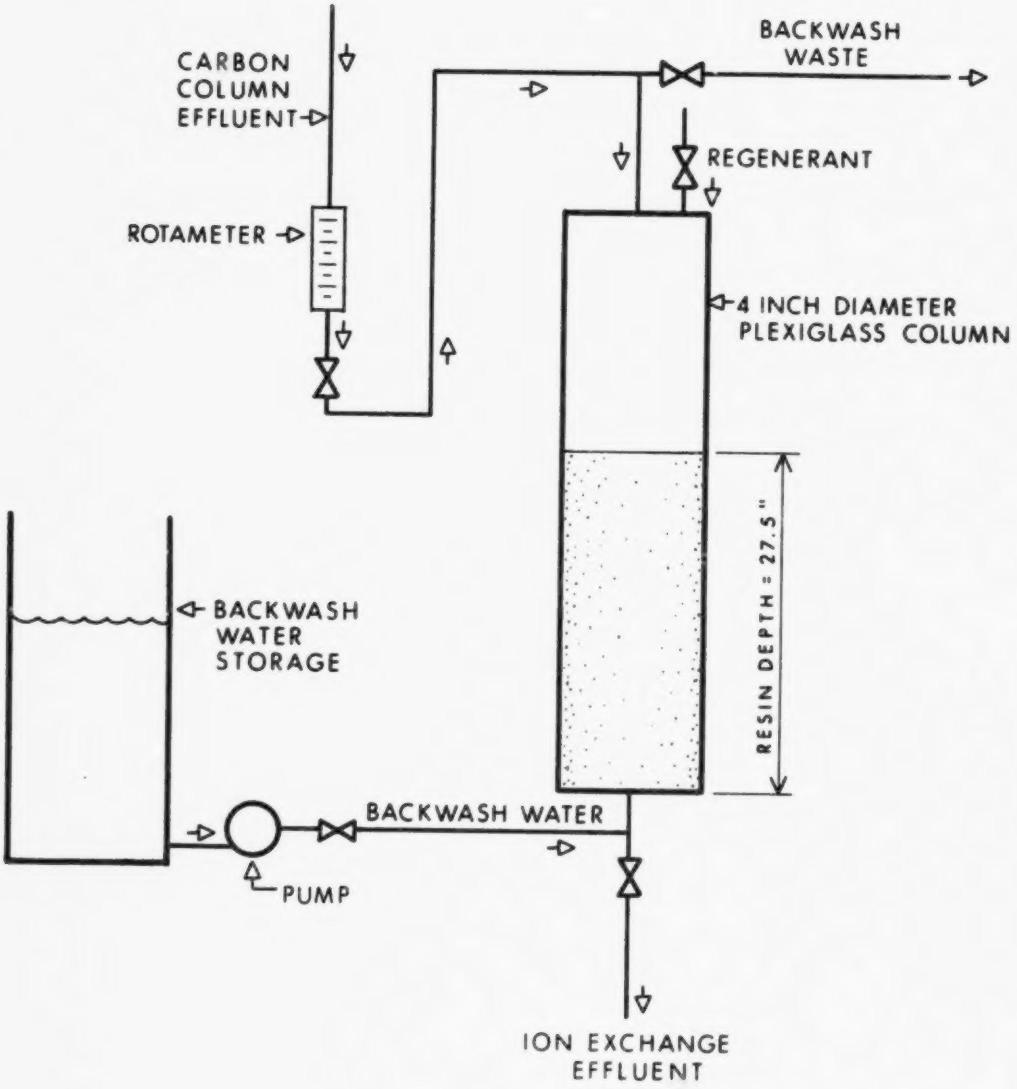

CARBON
COLUMN
EFFLUENT
ROTAMETER
BACKWASH
WASTE
REGENERANT
4 INCH DIAMETER
PLEXIGLASS COLUMN
RESIN DEPTH = 27.5"
BACKWASH
WATER
STORAGE
BACKWASH WATER
PUMP
ION EXCHANGE
EFFLUENT

FIGURE 80

SCHEMATIC DIAGRAM OF
ION EXCHANGE PILOT PLANT

The manufacturer's recommendations for operating conditions have been followed in these preliminary studies. A service flow rate of 3 gpm per cubic foot of resin was used. Following breakthrough of the ammonia-nitrogen to 1 mg/l, the bed was backwashed with carbon column effluent at a rate of about 5 gpm per square foot for 5 minutes. Because of the economic advantages and ease in handling, sodium chloride rather than acid was used as a regenerant in these studies. The manufacturer recommends that 5 to 20 pounds of salt per cubic foot of resin should be applied in a 10- to 20-percent salt solution. In these preliminary studies, 10 pounds of salt per cubic foot were applied in a 10-percent salt solution. The salt solution was applied to the bed at a rate of 0.2 gpm per cubic foot (corresponding to 60 minutes contact time), and the bed was then rinsed with 40 gallons of carbon column effluent per cubic foot of resin at a rate of 1 gpm per cubic foot of resin.

Grab samples of the ion exchange column influent and effluent were collected hourly and analyzed for ammonia-nitrogen.

Experimental Results

Table 22 summarizes the ion exchange data collected during these studies. Only three runs were completed. Entrapped air in the resin forced one run to be discontinued. After several runs under the operating conditions described above, it was planned to vary the amount of salt used for regeneration to determine the minimum regenerant conditions. Unfortunately, continued mechanical and operational problems with the secondary plant prevented any further runs during this grant period.

It is apparent that ion exchange treatment of the tertiary effluent would provide an effluent with an ammonia-nitrogen content of less than 1 mg/l. An average of 365 bed volumes of carbon column effluent were treated prior to breakthrough (1 mg/l ammonia-nitrogen).

Feasibility of Process

Based on the same operating procedure used in the pilot ion exchange unit, a bed containing 1,280 cubic feet of resin would be required to treat 5 mgd of tertiary effluent. About 8.75 tons of salt per day would be required for regeneration of the resin. Based upon a resin cost of $20.50 per cubic foot, amortization of the resin cost over 20 years at 4 percent, and a salt cost of $25.40 per ton, the cost for resin and salt alone would be $46 per million gallons for a flow of 5 mgd. The salt cost constitutes $44.50 of this total. Thus, if a significantly smaller amount of salt is found to satisfactorily regenerate the resin, the cost of ion exchange treatment could be reduced significantly. The salt cost could also be reduced if a portion of the regenerant could be reused. The minimum salt requirement and the feasibility of regenerant reuse are two aspects which merit further study.

Table 22

ION EXCHANGE DATA

Bed Volumes Passed	NH3-N, mg/l Influent	 Effluent	Percent Removal
Run One			
24	27.5	4.2	85
47		4.2	85
73		3.7	87
103		2.9	90
139		2.0	93
167	23.5	1.6	93
193		1.3	94
210		0.9	96
234		1.1	95
252		0.8	97
270		0.9	96
288		1.1	95
306		1.2	95
325		1.4	94
348		16.0	26
Run Two			
23	21.5	2.8	87
43	21.0	2.2	90
87	20.2	1.6	92
118	22.0	1.1	95
148	20.5	0.7	96.5
200	21.0	0.6	97.5
224	21.0	0.3	98.5
254	21.0	0.1	99.5
272	21.0	0.2	99
296	21.0	0.1	99.5
321	21.0	0.3	98.5
345	21.0	0.3	98.5
369	21.0	0.4	98
394	21.0	0.7	96.5
530	21.0	6.6	69
555	18.0	8.8	51
585	18.0	13.0	28
600	18.0	15.0	17
623	18.0	17.0	6
Run Three			
24	22.0		
45		4.0	82
121		3.4	85
170		2.0	91
217		1.4	94
266		0.9	96
294	30	0.8	97
318	26	0.5	98
342	27	0.5	98
366	28.5	0.6	98
384	27	0.8	97
408	26	1.2	95
432	30	2.0	93
456	28	3.4	88
477		4.7	80
510		8.8	68
530		10.6	62
553		12.8	54
580		16.8	40
595		18.8	33
610		20.8	26
635		23.0	18
660		26.0	7

Even if the regenerant requirement could be reduced, the problem of brine disposal at Lake Tahoe remains economically difficult. It is unlikely that the brine could be disposed of within the Lake Tahoe Basin. Without reuse of the regenerant, the brine volume would be about 0.5 percent of the flow treated or about 25,000 gallons per day for a 5-mgd plant. The cost to truck this volume of brine to a disposal site outside of the basin would be prohibitive. The feasibility of regenerant reuse, the minimum salt requirement, and the feasibility of using a more concentrated salt solution should be determined to better evaluate the brine disposal problem, which is of special importance in the Lake Tahoe Basin.

AMMONIA STRIPPING

Theory

Ammonium ions exist in equilibrium with ammonia and hydrogen ions in the waste water as shown by the equation:

$$NH_4^+ \rightleftharpoons NH_3\uparrow + H^+$$

As the pH of the waste water is increased above 7, the equilibrium is displaced to the right so that ammonia may be liberated as a gas. Theory indicates that the equilibrium shifts further to the right, as the pH is increased from 7.0 to 11.5. Figure 81 illustrates the theoretical relation between pH and percent ammonia available for removal by stripping at 77°F.(38) The curve on Figure 81 shifts to the left as the temperature decreases.

Following the increase of the pH of the waste water to a high level, agitation of the waste water is necessary to liberate the ammonia gas. The air-water contact provided by the drop formation and reformation in a stripping tower has been found to be a very efficient method of removing ammonia gas from solution.(43) The hydrogen ion liberated from the ammonium ion will result in a decrease in pH as the ammonia is driven off as a gas.

About 90 percent of the nitrogen in the raw sewage at South Tahoe P. U. D. is in the form of ammonia or compounds from which ammonia is readily formed. If proper environmental conditions are provided in the secondary treatment process, the ammonia will be converted to nitrate. However, this nitrification process can be eliminated by maintaining a relatively high organic loading on the secondary process. It is cheaper and easier to maintain the nitrogen in the form of ammonia than it is to convert it to nitrate. Thus, the removal of nitrogen in the form of ammonia offers some economic and operational advantages over the removal of nitrate-nitrogen.

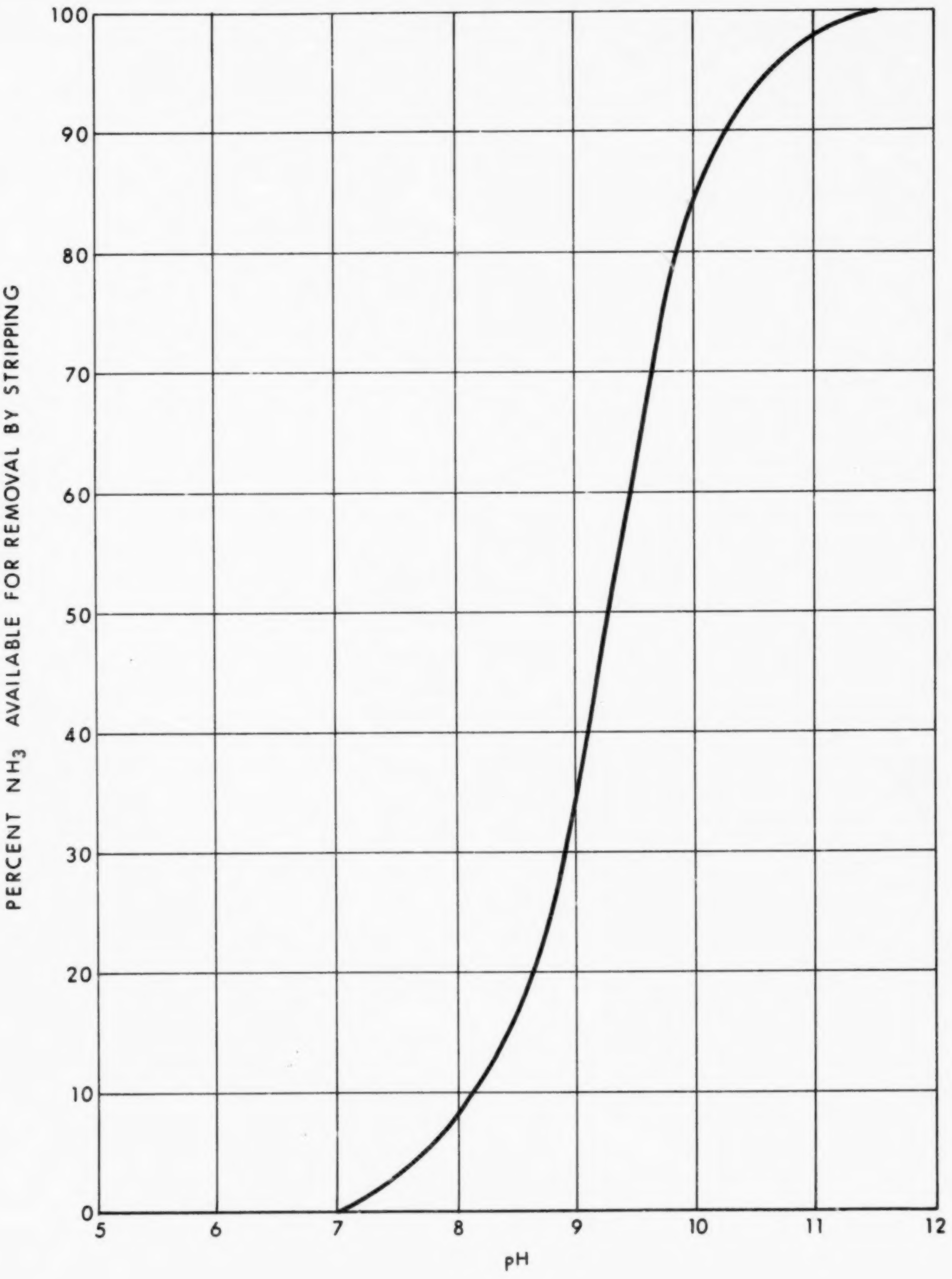


FIGURE 81

PERCENT AMMONIA AVAILABLE FOR REMOVAL
BY STRIPPING AT 77° F. AS A FUNCTION OF pH

Literature Review

Very little work has been done with ammonia stripping with concentrations as dilute as those found in sewage.

Kuhn(44) studied removal of nitrogen from municipal waste water by ammonia stripping. He found that ammonia-nitrogen removal increased as the pH of the waste water was increased from 8 to 11. No significant increase in the removal of ammonia-nitrogen was obtained by increasing the pH from 11 to 12. Ammonia-nitrogen removals were found to increase as the air/liquid loading was increased from 40 to 447 cubic feet of air per gallon of waste water. A tower packed with 1/2-inch Raschig rings and loaded at a rate of 0.3 gpm per square foot was used in these studies. The optimum ammonia-nitrogen removal of 92 percent was obtained at a pH of 11 with an air/liquid loading of about 500 cubic feet per gallon and a tower depth of 7 feet.

Prather has reported several studies(43,45,46) in which ammonia was removed from petroleum refinery waste water by stripping with air. The concentrations of ammonia-nitrogen in the untreated waste water averaged slightly more than 100 mg/l. When 300 cubic feet of air were applied per gallon of waste water, the ammonia removal was found to be 85 percent at a pH of 10.5 and 34 percent at a pH of 9.4.(46) In another study(43) in which the waste water was passed through a closely packed aeration tower with 480 cubic feet of air supplied per gallon, ammonia-nitrogen removal by air stripping was found to be very effective (more than 95 percent removal) at any pH level above 9.0. When the pH fell below 9.0, the ammonia-nitrogen removal decreased sharply. The removal fell to 91 percent at a pH of 8.9 and to 58 percent at a pH of 8.8.

Experimental Procedures

A small air-conditioning tower designed for home use (Marley Aquatower Model 4411), packed with redwood slats, was used as a stripping tower in this study. The packing of this tower is compact, having a depth of 2 feet and an area of 5 sq ft. The tower is equipped with a fan which draws 1,520 cfm of air through the tower. This type of tower was used to evaluate the effect of aeration following pH adjustment, recognizing that this type of tower packing does not provide optimum air-water contact.

In each of several runs, the pH of 22 gallons of tertiary effluent was adjusted with sodium hydroxide. This effluent was then recycled through the tower (Figure 82) with frequent measurements of the ammonia-nitrogen content of the tower effluent being made. The pH of tertiary effluent (tower influent) was adjusted to values ranging from 8.0 to 10.8, and the flow to the tower was varied from 0.5 gpm to 4 gpm (0.1 to 0.8 gpm/sq ft).

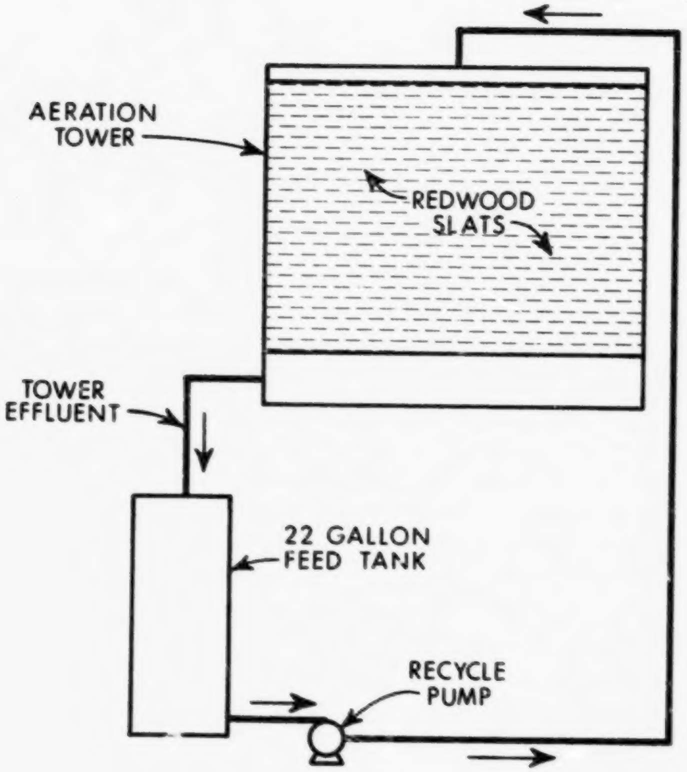

AERATION TOWER
REDWOOD SLATS
TOWER EFFLUENT
22 GALLON FEED TANK
RECYCLE PUMP

FIGURE 82

SCHEMATIC DIAGRAM OF AMMONIA STRIPPING PILOT PLANT

Experimental Results

The efficiency of ammonia-nitrogen removal in the particular aeration tower used in these studies at various pH values is shown on Figures 83. Optimum ammonia removals at pH values above 9.0 were obtained at an aeration time of about 0.5 minute, or, with this particular tower, at an air/liquid loading of about 750 cubic feet per gallon. Kuhn(44) and Prather(43) have found that an air/liquid loading of 450 cubic feet per gallon is adequate for ammonia stripping in towers which provide efficient air-water contact. With the tower used in this study, the optimum aeration time corresponded to a tower depth of about 20 feet.

Approximately 94 percent of the nitrogen in the tertiary effluent is in the form of ammonia-nitrogen. Thus, the 98 percent ammonia-nitrogen removal obtained at a pH of 10.8 corresponds to a removal of about 92 percent of all nitrogen present in the tertiary effluent. At a pH of 9.7, 93 percent of the ammonia-nitrogen, or 87 percent of all the nitrogen in the tertiary effluent were removed. Organic nitrogen was not affected by passage through the tower.

Table 23 summarizes the ammonia-nitrogen content of the aeration tower influent and effluent at various pH values. The values shown for tower effluent in Table 23 are those occurring at the end of each run shown on Figure 83.

Table 23

AMMONIA-NITROGEN CONTENT OF AERATION TOWER
INFLUENT AND EFFLUENT

Tower Influent			Tower Effluent			
pH	Temp. (°C)	NH_3-N (mg/l)	pH	Temp. (°C)	NH_3-N (mg/l)	Percent NH_3-N Removal
8.0	22	28.8	8.0	16	18.2	37
8.6	21	26.6	8.1	17	15.8	41
9.3	23	26.2	8.6	18	5.3	80
9.7	21	30.0	9.1	16	2.1	93
10.8	22	25.5	10.1	17	0.6	98

Figure 84 shows a flow sheet of possible potential modifications to the existing tertiary treatment plant to provide both phosphate and nitrogen removal using lime as the coagulant. A savings in cost of chemicals might be realized if lime is used as the coagulant, rather than alum. The lime would remove the phosphates present in the secondary effluent by precipitation of calcium phosphate and also raise the pH sufficiently so that ammonia stripping could be accomplished. It should be noted that pH adjustment (by recarbonation) would be accomplished prior to passing the water through the activated carbon. A pH of 11 would have an adverse effect on the carbon adsorption process.

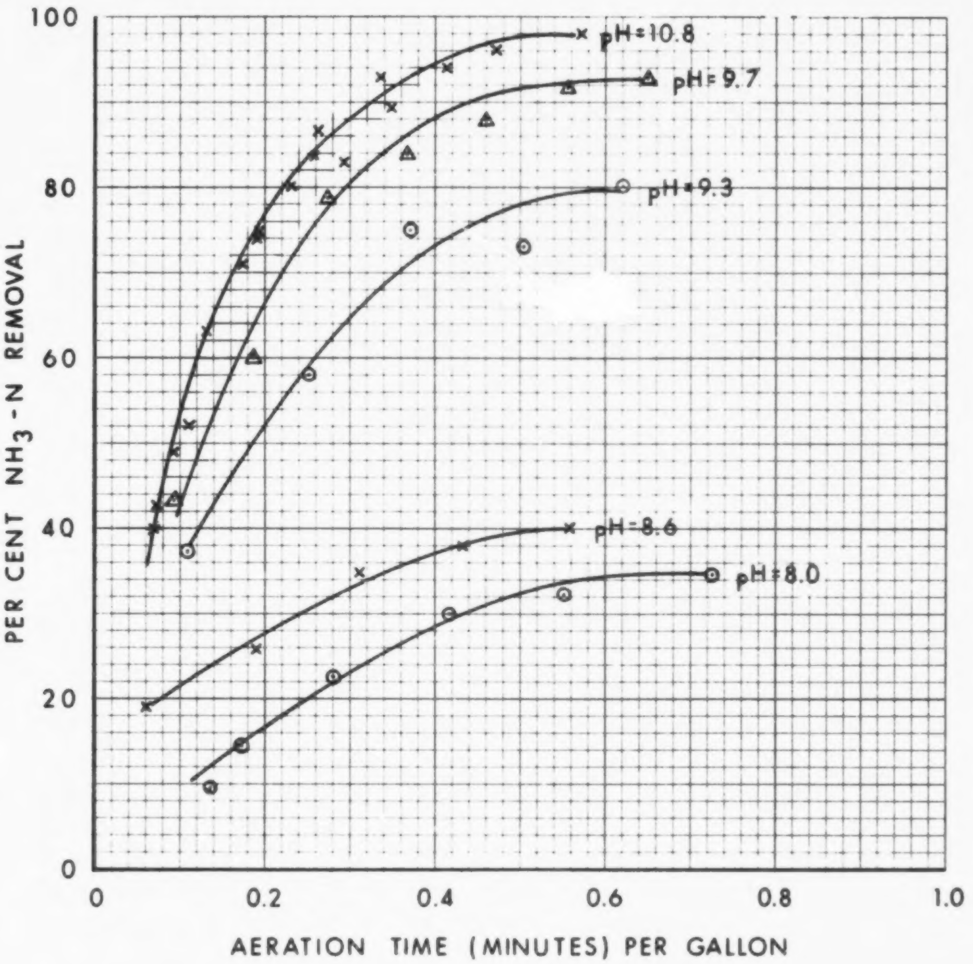


FIGURE 83
PERCENT AMMONIA REMOVAL
VERSUS AERATION TIME

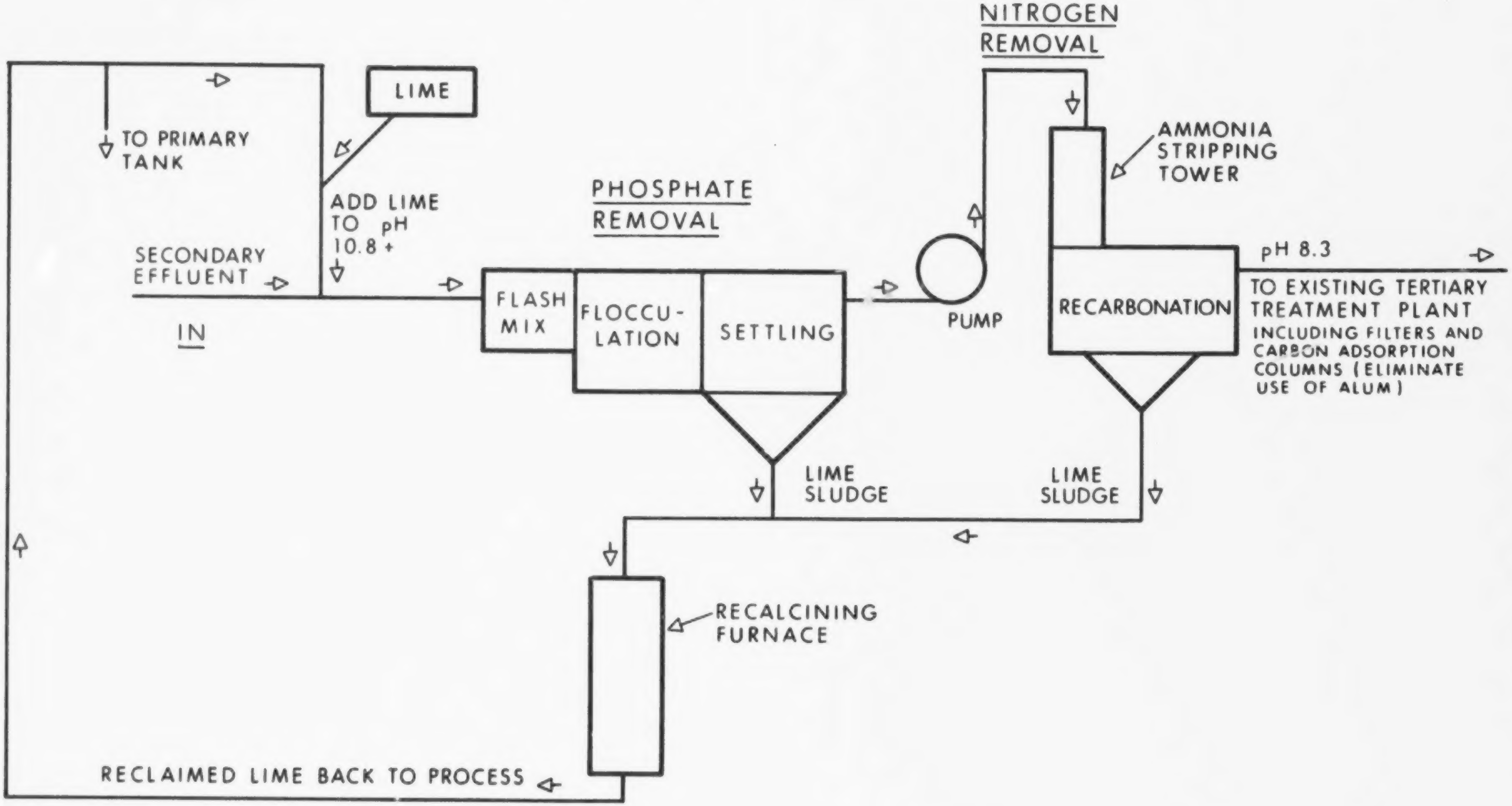

NITROGEN REMOVAL
TO PRIMARY TANK
LIME
AMMONIA STRIPPING TOWER
ADD LIME TO pH 10.8 +
PHOSPHATE REMOVAL
SECONDARY EFFLUENT
IN
FLASH MIX
FLOCCU-LATION
SETTLING
PUMP
RECARBONATION
pH 8.3
TO EXISTING TERTIARY TREATMENT PLANT
INCLUDING FILTERS AND CARBON ADSORPTION COLUMNS (ELIMINATE USE OF ALUM)
LIME SLUDGE
LIME SLUDGE
RECALCINING FURNACE
RECLAIMED LIME BACK TO PROCESS

FIGURE 84

PROPOSED FLOW SHEET FOR REMOVAL OF PHOSPHATE AND NITROGEN BY USE AND RECOVERY OF LIME

Concurrent laboratory and pilot plant tests were made to determine the feasibility of using lime as the coagulant in the tertiary process and for pH adjustment. The results of this study were presented in the chapter on coagulant recovery. (Chapter VI)

Review of Nitrogen Removal Methods

At the conclusion of the initial nitrogen removal studies, an evaluation of the various methods was prepared. Nitrogen removal by ammonia stripping was found to have several advantages over the other methods of nitrogen removal studied:

1. The process provides a positive control of the effluent nitrogen at all times. The process does not suffer the wide fluctuations in efficiency as shown by the denitrification process.

2. A lower effluent nitrogen content can be maintained than with the denitrification or ion exchange processes.

3. There is no need to convert the ammonia-nitrogen to nitrate-nitrogen, a requirement of the denitrification process. This will simplify operation and lower the cost of secondary treatment.

4. The nitrogen is removed in gaseous form, and there are no liquid or solid nitrogenous wastes requiring final disposal. Operation would be much simpler than that required by an ion exchange installation; i.e., there is no regeneration process which involves handling of salt or acid solutions, backwashing and rinsing the resin, and continuous monitoring of the final effluent for ammonia-nitrogen. In addition, the disposal of the waste brine in the Lake Tahoe Basin would be prohibited, requiring that it be exported from the basin.

Pilot Ammonia Stripping Tower

It was felt a larger pilot stripping tower was essential to further refine the possibilities and design criteria for a plant-scale tower. In cooperation with the Marley Company, Kansas City, Missouri, a pilot-scale tower with a larger capacity was designed and constructed.

Experimental Procedures and Results. Figure 85 shows a schematic of the ammonia stripping tower. The pilot tower is 25 feet high, 6 feet wide, and 4 feet in depth. A photograph of the tower is shown in Figure 86. The tower is completely encased except for the air inlet at the bottom and the exit through the fan near the top. The tower contains six sections of close fill, having a cross sectional area of 7.5 sq ft. The fill which consisted of redwood slats

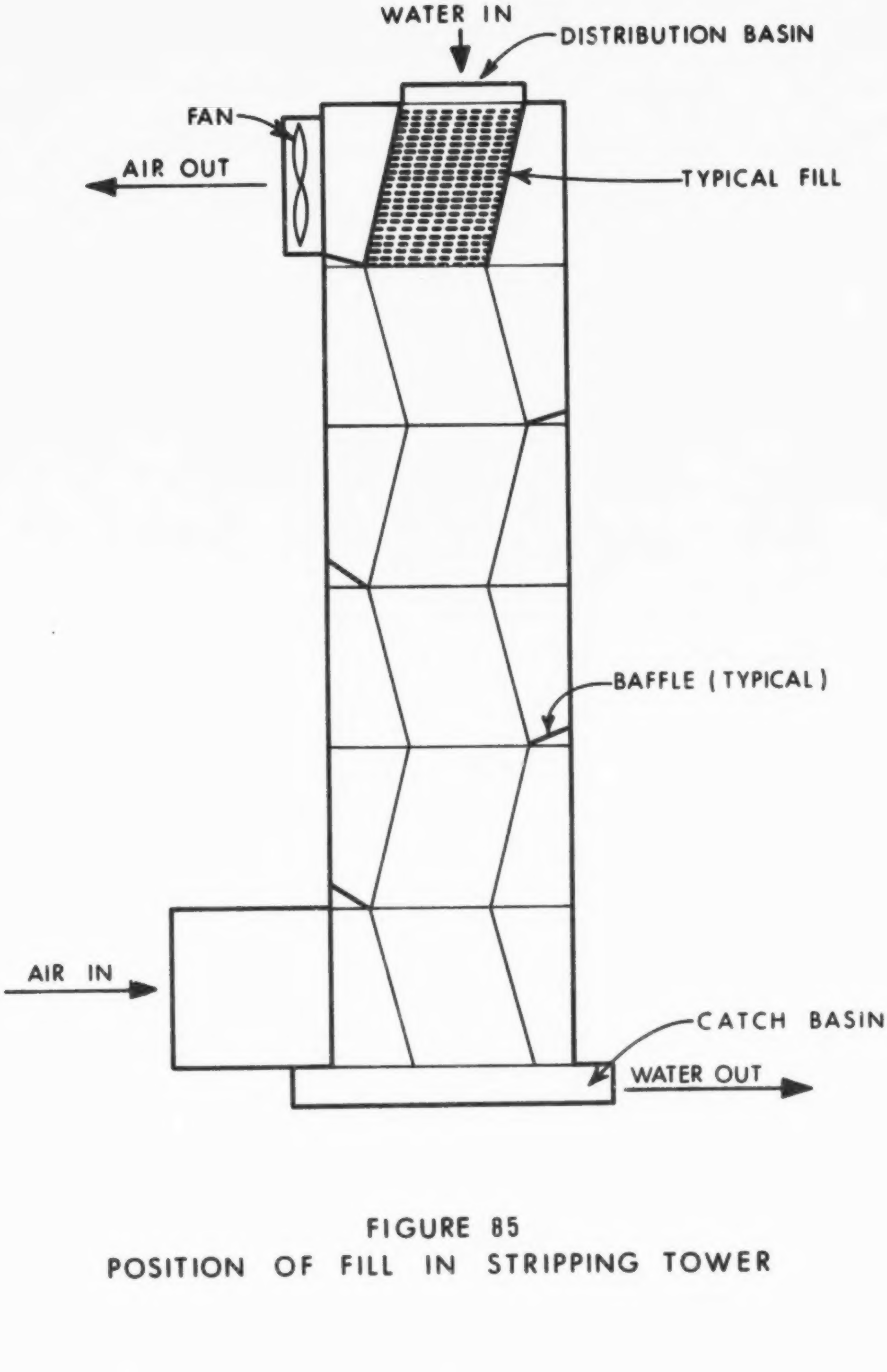

WATER IN
DISTRIBUTION BASIN
FAN
AIR OUT
TYPICAL FILL
BAFFLE (TYPICAL)
AIR IN
CATCH BASIN
WATER OUT

FIGURE 85
POSITION OF FILL IN STRIPPING TOWER

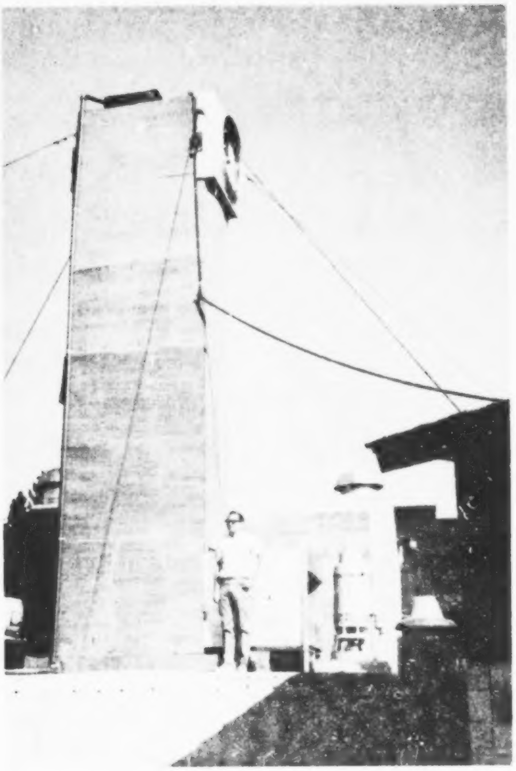


FIGURE 86
PILOT SCALE NITROGEN STRIPPING TOWER
S.O. TAHOE P.U.D. WATER RECLAMATION PLANT

3/8-inch-thick by 1-1/2 inches wide, was spaced 2 inches on center horizontally and 1-1/2 inches on center vertically. The sections of fill were placed one above the other in a diagonal position, and baffles were provided so that air entering the bottom of the tower would be directed almost horizontally through each section of fill before leaving the tower through the fan at the top of the tower. The fan is 42 inches in diameter and is driven by a 5 hp motor.

Secondary effluent was used as a source of water for the ammonia stripping tower. The flow to the tower was measured by a water meter and the flow rate controlled by a gate valve. After metering, the pH was raised to a value of 11.5, the pH value which would be attained in the full-scale plant, using lime as a coagulant by injecting caustic soda into the pressure line feeding the tower. This water was discharged into a distribution basin at the top of the tower, which has an area of 10 sq ft. There are 24-1/2-inch-diameter holes evenly spaced over the distribution box floor.

Samples for ammonia analyses were obtained from sampling trays located at various depths in the tower. The surface-loading rate and air supply were varied to study the effects of these two factors on the efficiency of ammonia removal for different depths of tower packing.

For a given supply of air per gallon of waste water treated (Figure 87), the efficiency of ammonia stripping increases with increased packing depths. This appears to be due to the greater number of droplet formations and the resulting increased air-water contact achieved with increased packing depths. The efficiencies of the 20- and 24-foot packing depths were essentially the same for ammonia removals of up to 90 percent. Removals above 90 percent were difficult to achieve with the 20-foot packing, even with increased quantities of air, while it was possible to obtain ammonia removals of 98 percent with the 24-foot packing.

There was little difference between the efficiency of the 20- and 24-foot towers at flow rates up to 7 gpm/sq ft, as shown by Figure 88. The rapid decrease in removals for loadings greater than 3 gpm/sq ft was due to the observed sheeting of water in the tower. Sheeting of the water decreases the number of droplet formations and consequently reduces the efficiency of ammonia stripping.

Figure 89 shows that the spacing of the tower fill is very important in determining the air requirements for ammonia stripping. The 1-1/2-inch by 2-inch packing has two and two thirds more slats for droplet rupture and formation than does the 4-inch by 4-inch packing. The data indicate that droplet rupture is a key factor in obtaining high ammonia removal.

Figure 90 shows the efficiency of ammonia removals decreases more rapidly with increased surface-loading rates with the 4-inch by 4-inch packing than with the 1-1/2-inch by 2-inch packing. The degree of sheeting was observed to be less in the 1-1/2-inch by 2-inch packing.

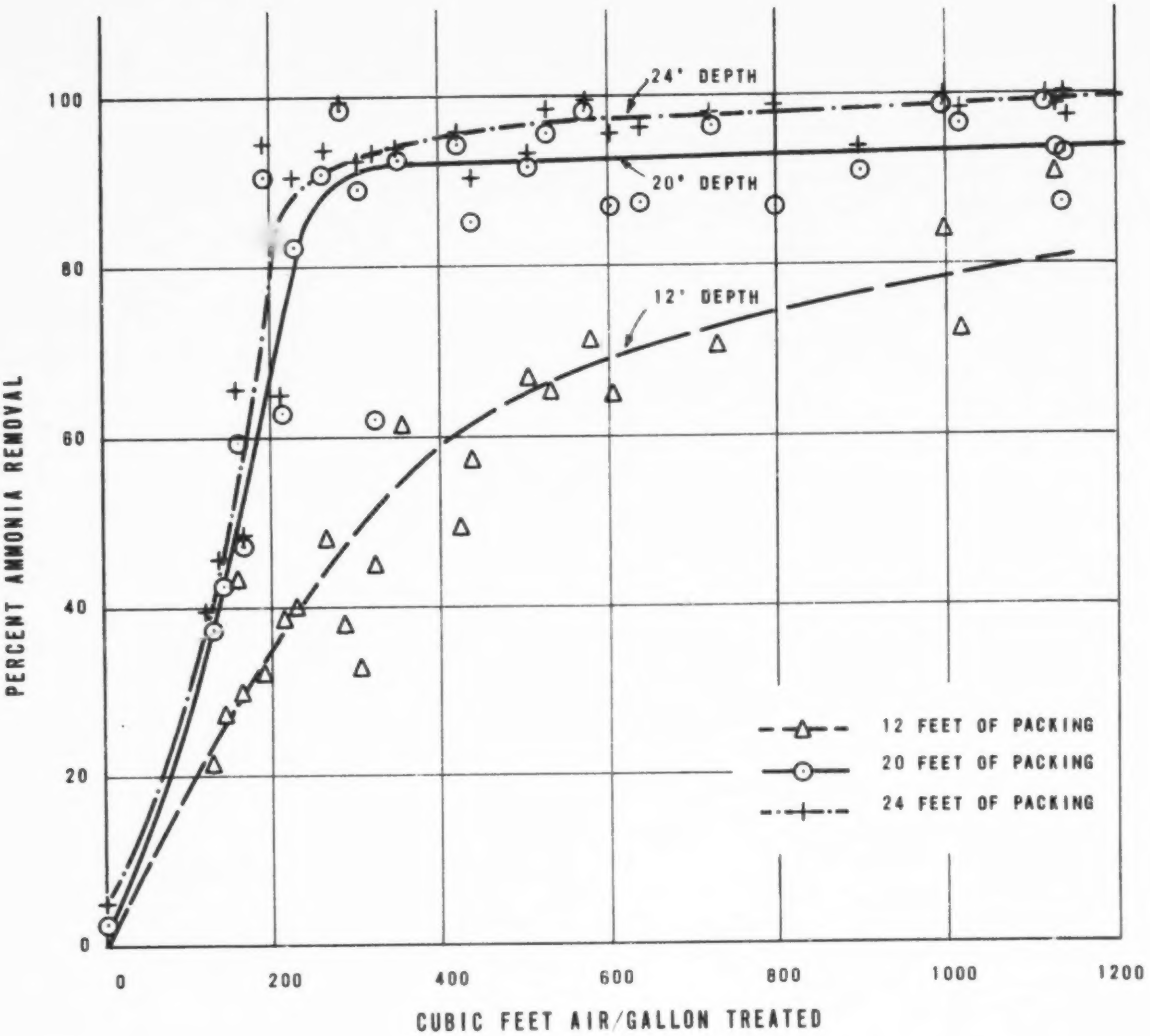


FIGURE 87

PERCENT AMMONIA REMOVAL VS. CUBIC FEET OF AIR PER GALLON WASTEWATER TREATED FOR VARIOUS DEPTHS OF PACKING

C4125.0

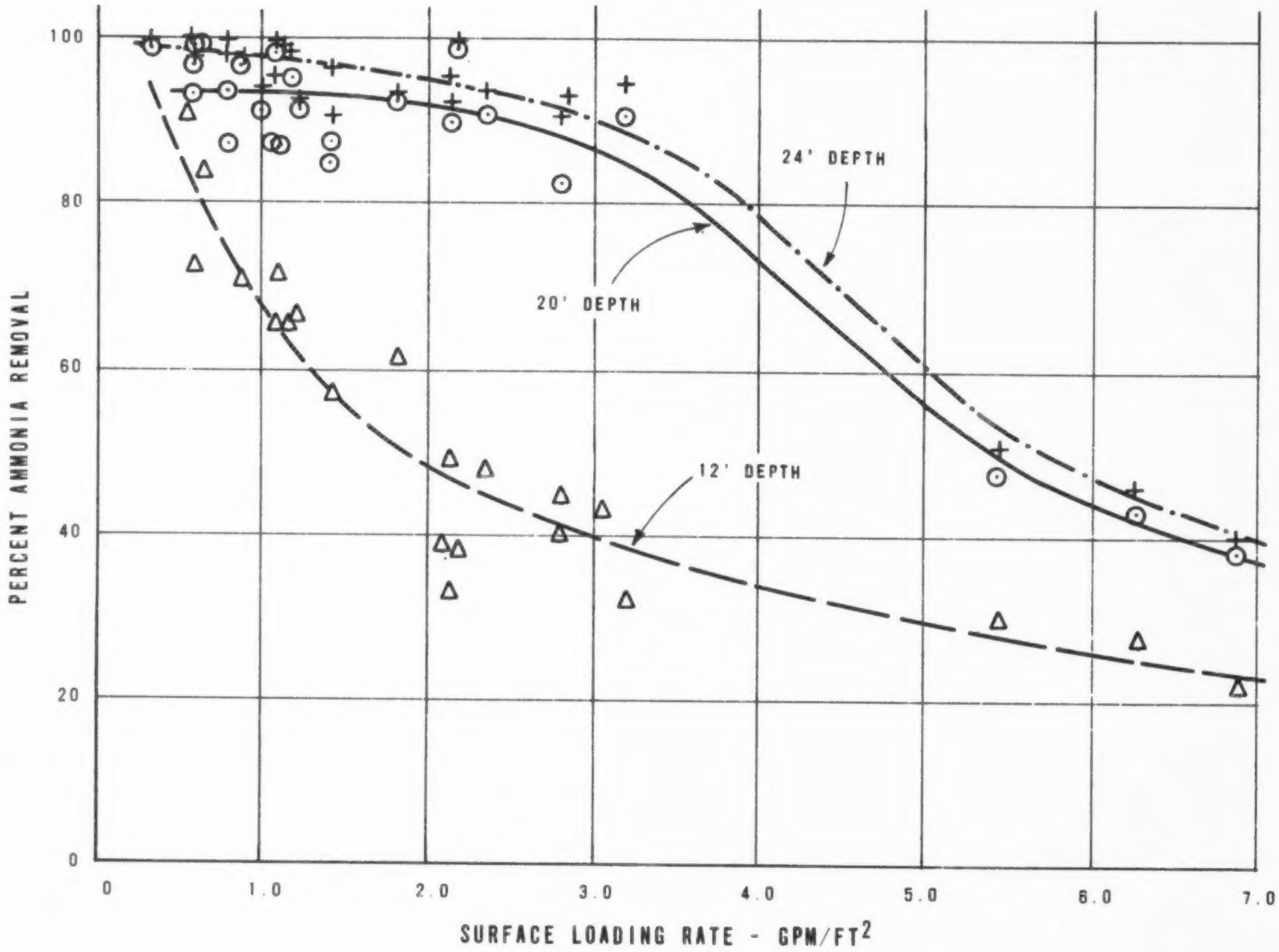


FIGURE 88

PERCENT AMMONIA REMOVAL VS. SURFACE LOADING RATE FOR VARIOUS DEPTHS OF PACKING

C4125.0

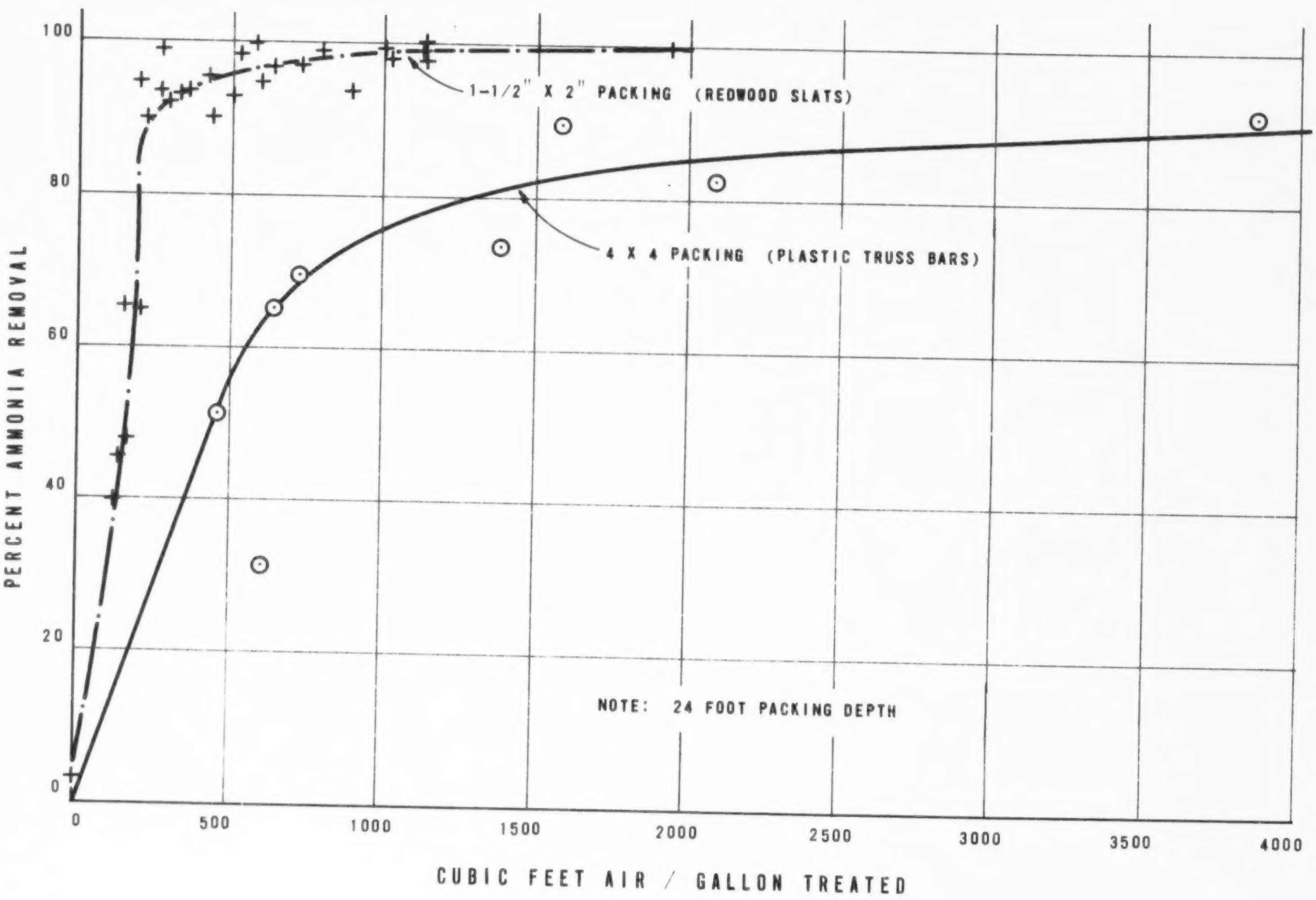


FIGURE 89

EFFECT OF PACKING SPACING ON AIR REQUIREMENTS AND EFFICIENCY OF AMMONIA STRIPPING

C4125.0

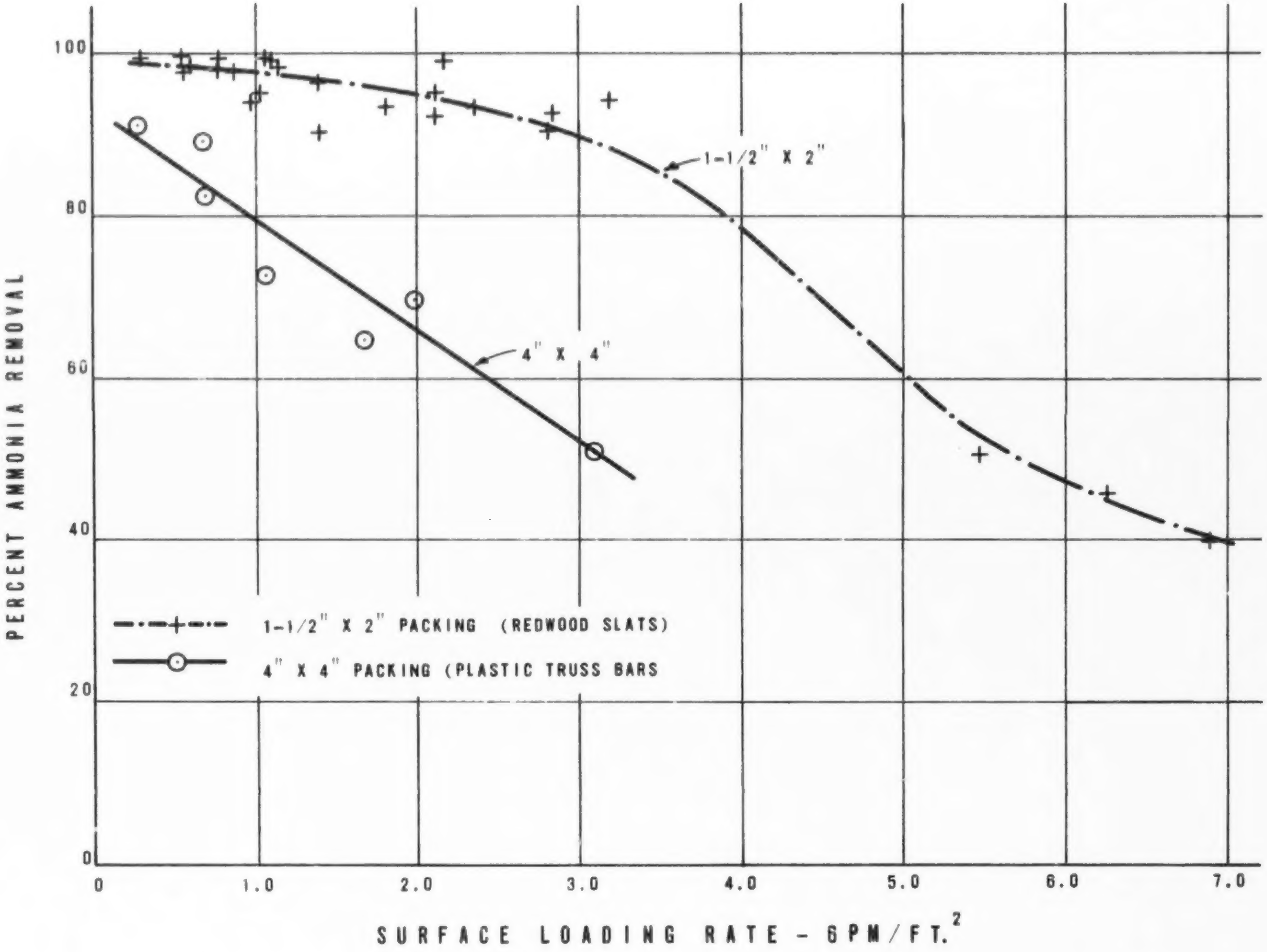


FIGURE 90

EFFECT OF PACKING SPACING
ON SURFACE LOADING RATE AND EFFICIENCY OF AMMONIA STRIPPING

C4125.0

Table 24 summarizes the air and hydraulic loading requirements for various ammonia-nitrogen removals using 20- and 24-foot towers with a 1-1/2-inch by 2-inch packing. For the same ammonia-nitrogen removal, it is possible to use less air and higher surface-loading rates with the 24-foot tower.

It can be seen that 90 percent ammonia-nitrogen removal is possible with approximately 250 cubic feet of air per gallon treated, while the quantities of air required for higher removals increase rapidly (400 cu ft/gallon for 95 percent removal and 800 cu ft/gallon for 98 percent removal).

Effect of Temperature on Ammonia Stripping. The above data and results were obtained during summer operating conditions when the temperature of the waste water being treated was always above 20°C and ambient temperatures were high. The tower was later operated during winter conditions in conjunction with a lime clarification unit described later in this report. At that time, it was observed that the colder air and water temperatures had a pronounced effect on the ammonia stripping efficiency.

According to theory, the rate of ammonia stripping should be a function of the difference in partial pressure of the ammonia in the liquid phase and the gaseous phase.

$$-\frac{dC}{dt} \propto \left(P_{\text{liquid}} - P_{\text{gas}}\right)$$

Ammonia partial pressure data could not be located which dealt with ammonia concentrations as low as those involved in this process. The data did indicate, however, that a 10°C decrease in process water temperature would decrease the difference in partial pressures by about 40 percent.(47) This, theoretically, decreases the driving force for ammonia stripping by this amount.

Figure 91 presents ammonia removal data collected on one particular day with all operating conditions except temperature held constant. Also included, for comparative purposes, are stripping data collected under similar circumstances during the previous summer. This figure illustrates quite well the decrease in efficiency resultant from a decrease in operating temperature.

The removal efficiency of the tower is substantially reduced as the temperature falls below 20°C, but a lower limit of removal efficiency is apparently reached as the temperature approaches 0°C. A change in effluent temperature from 3°C to 9°C improved ammonia removal by only 5 percent, while a 3°C temperature differential between 9°C and 12°C resulted in an 8-percent increase in ammonia removal. Of course, the percent ammonia removal is dependent upon many factors. Data collected from operation of the tower during winter conditions indicate the average lower limit of the process will be in the range of 50 to 60 percent ammonia removal.

Table 24

AIR- AND SURFACE-LOADING REQUIREMENTS FOR 20- AND 24-FOOT TOWERS FOR VARIOUS AMMONIA-NITROGEN REMOVALS BY AMMONIA STRIPPING (1)

Percent Ammonia-Nitrogen Removal	20-Ft Depth, 1-1/2" x 2" Packing		24-Ft Depth, 1-1/2" x 2" Packing	
	Ft^3 Air/Gallon	Hydraulic Loading gpm/ft^2	Ft^3 Air/Gallon	Hydraulic Loading gpm/ft^2
80	230	3.6	200	3.9
85	240	3.2	210	3.55
90	280	2.5	250	3.0
95	—	—	400	2.0
98	—	—	800	0.8

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(1) Effluent water temperature ≈ 20°C.

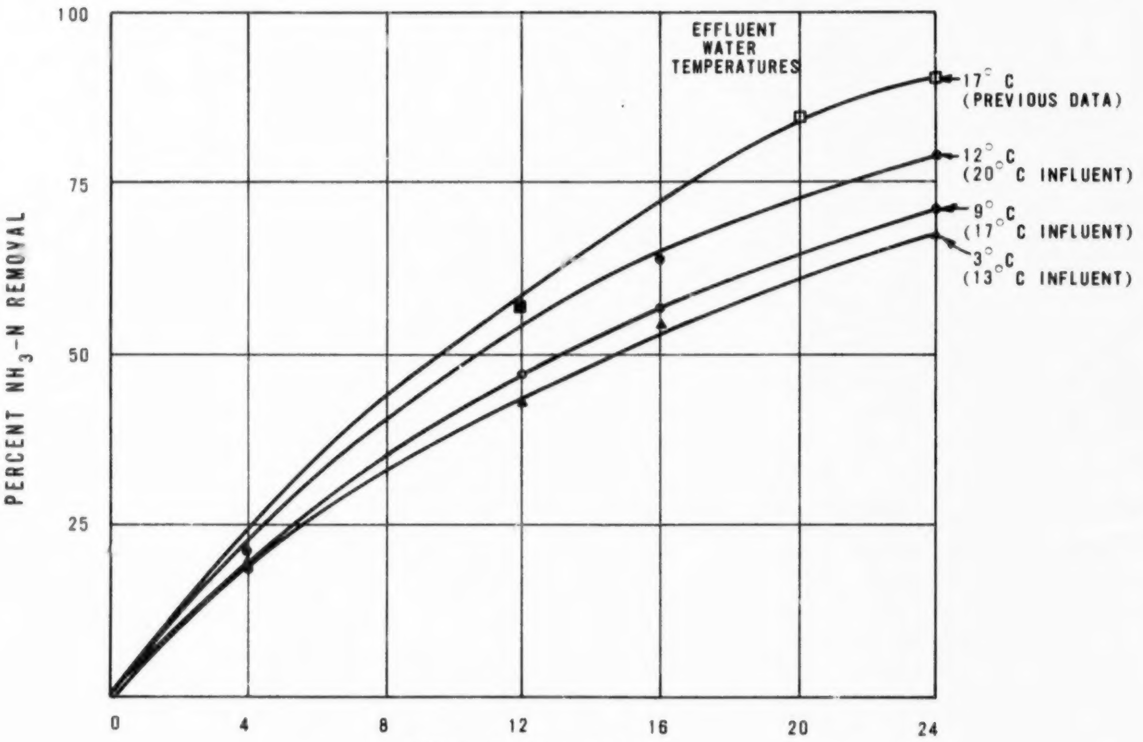


FIGURE 91

EFFECT OF WATER TEMPERATURE ON AMMONIA STRIPPING
(OPERATING CONDITIONS - 2.0 GPM/FT^2, 480 FT^3 AIR/GALLON)

It appears that a taller tower, or provisions for recirculation, would be beneficial under winter conditions. After 24 feet of packing contact, a limit for ammonia removal had not been reached at any temperature level. The rate of removal would diminish with decreasing temperature and ammonia concentration. Therefore, the effect of providing additional packing contact must still be investigated.

Another factor resultant from winter operating conditions is ice formation on the tower packing. As the temperature of the water passing through the tower approached 0°C, icicles formed on the packing to the point where droplet formation was disrupted. This limiting condition might be minimized by:

1. Flooding the packing under no-draft conditions, or
2. Reversing the air flow

Estimated Cost of Ammonia Stripping in Conjunction with Lime Coagulation. An estimate was prepared for the cost of removing ammonia-nitrogen by air stripping after using lime as a coagulant. The process schematic envisioned for the accomplishment of this nutrient removal is presented in Figure 92.

Table 25 presents the cost for various degrees of ammonia stripping. The table was originally compiled from data collected during summer operating conditions. This indicated that for a total cost of $13.85 per million gallons, 95 percent removal could be obtained. Later data indicate the same operating expense will yield only approximately 60 percent ammonia removal during the winter months. The requirements for increased removal under these conditions have not been established.

CHLORAMINE ADSORPTION AND BREAKPOINT CHLORINATION

Theory

When chlorine is added to water containing ammonia-nitrogen, the ammonia reacts with the chlorine (hypochlorous acid) to form various chloramines. The type of chloramine formed is dependent upon the ratio of the concentrations of chlorine to ammonia, the pH of the reaction, and the reaction time.(48,49)

$$NH_3 + HOCl \longrightarrow NH_2Cl + H_2O \quad \text{monochloramine}$$

$$NH_3 + 2HOCl \longrightarrow NHCl_2 + H_2O \quad \text{dichloromine}$$

$$NH_3 + 3HOCl \longrightarrow NCl_3 + 3H_2O \quad \text{trichloramine}$$

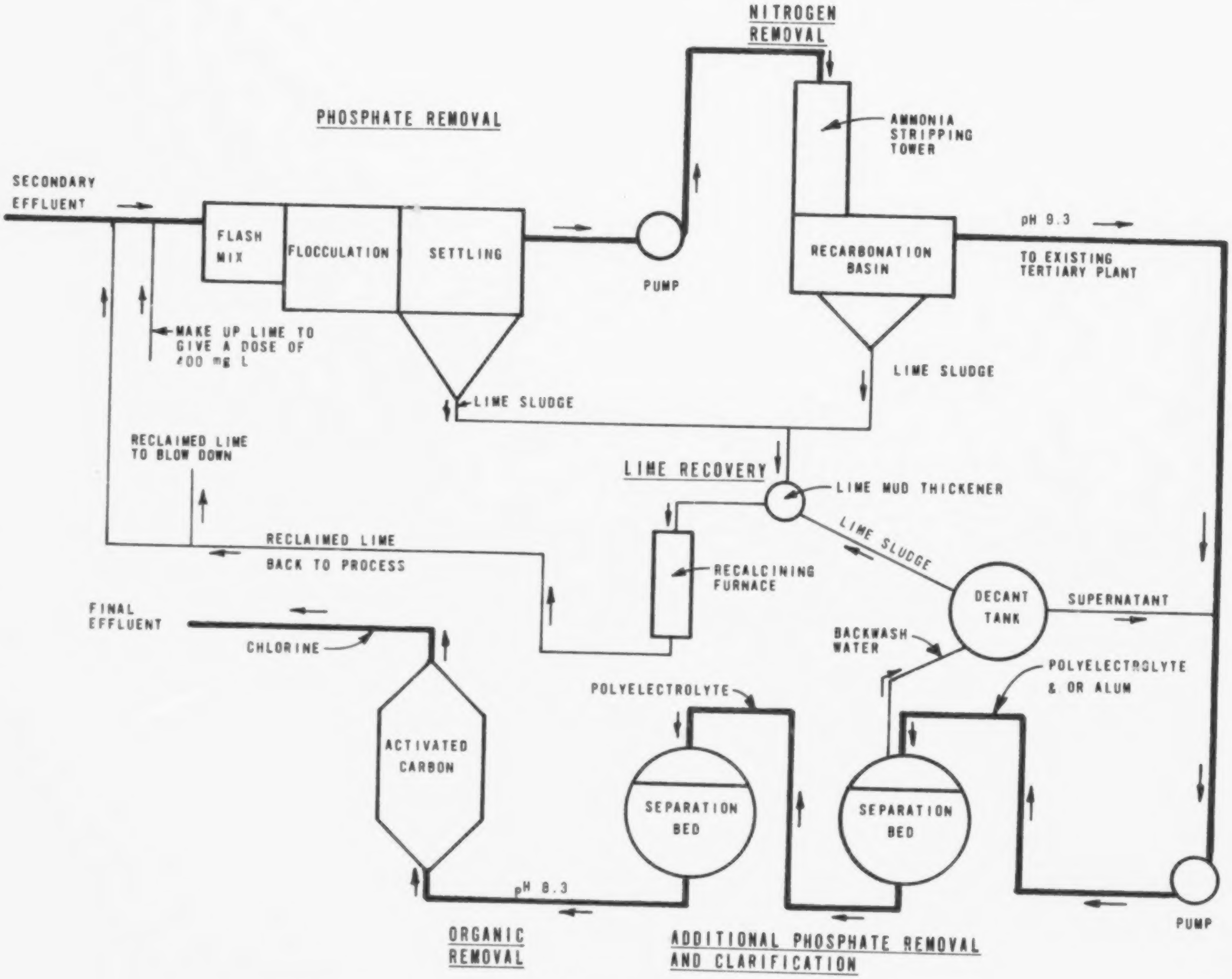

PHOSPHATE REMOVAL
NITROGEN REMOVAL
SECONDARY EFFLUENT
FLASH MIX
FLOCCULATION
SETTLING
PUMP
AMMONIA STRIPPING TOWER
RECARBONATION BASIN
pH 9.3
TO EXISTING TERTIARY PLANT
MAKE UP LIME TO GIVE A DOSE OF 400 mg L
LIME SLUDGE
LIME SLUDGE
RECLAIMED LIME TO BLOW DOWN
LIME RECOVERY
LIME MUD THICKENER
RECLAIMED LIME
BACK TO PROCESS
RECALCINING FURNACE
LIME SLUDGE
DECANT TANK
SUPERNATANT
FINAL EFFLUENT
CHLORINE
BACKWASH WATER
POLYELECTROLYTE
POLYELECTROLYTE & OR ALUM
ACTIVATED CARBON
SEPARATION BED
SEPARATION BED
pH 8.3
PUMP
ORGANIC REMOVAL
ADDITIONAL PHOSPHATE REMOVAL AND CLARIFICATION

FIGURE 92

PROPOSED PROCESS FOR PHOSPHATE, NITROGEN, ORGANIC REMOVAL AND RECOVERY OF COAGULANT

Table 25

ESTIMATED COST OF NITROGEN REMOVAL BY AMMONIA STRIPPING AT 7.5-MGD SOUTH TAHOE PUBLIC UTILITY DISTRICT WATER RECLAMATION PLANT

(Costs – Dollars/Million Gallons)

Percent Ammonia Removal		20-Foot Tower					24-Foot Tower				
		Capital	Power	Maintenance	Labor	Total	Capital	Power	Maintenance	Labor	Total
Summer	Winter										
80		2.70	4.00	0.20	2.00	8.90	2.85	3.75	0.20	2.00	8.80
90		3.65	4.70	0.20	2.00	10.55	3.60	4.40	0.20	2.00	10.20
95	60						5.10	6.50	0.25	2.00	13.85
98							11.80	10.90	0.30	2.00	25.00

Capital costs were amortized at 4 percent interest over a 20-year period. Power was based upon an energy charge of $0.008 per kw hour and a demand charge of $2.30 per kw hour per month. Maintenance costs were established at an annual cost of 3 percent of the equipment cost. Labor costs were based on $3.75 per hour. No additional chemical costs are necessary when the process is used in series with lime coagulation.

A preliminary study was undertaken to determine the ability of activated carbon to adsorb chloramines. It is well known that carbon is an excellent dechlorinating agent. What was not known was whether the ammonium ions of the chloramines would be removed along with the chlorine. If so, it might be more economical to accomplish about 90 percent ammonia removal by air stripping, and the remainder by chlorination and adsorption.

Experimental Procedures and Results

Lime-treated secondary effluent, recarbonated to a pH of 9.0 and containing approximately 15 mg/l $NH_3 - N$, was treated with sodium hypochlorite. The resulting chlorine residuals (below "breakpoint") were mostly in the form of monochloramine (NH_2Cl). Effluent samples with various dosages of chlorine as sodium hypochlorite were passed through a small carbon column with a 20-minute detention time. The following results were achieved (Table 26).

Table 26

RESULTS OF THE INVESTIGATION OF CHLORAMINE

Adsorption by Activated Carbon
(Pittsburgh Type SGL)

Sample	Cl_2 Dose (mg/l)	Cl_2 Residual (mg/l)	$NH_3 - N$, mg/l	
			Before Carbon	After Carbon
A	37	24	15.4	14.2
B	50	29	15.0	13.4
C	50	28	15.0	14.2
D	75	52	15.4	14.6
E	110	36	12.3	9.8
F	150	2 (free)	3.7	—

To investigate the possibility of dichloramines being adsorbed more readily than monochloramines, another sample was chlorinated and the pH adjusted to approximately 4.0. After 2 hours, the chlorine analysis indicated 16 mg/l free chlorine, 5 mg/l monochlorine, and 15 mg/l dichloramine. Ammonia analysis before and after carbon treatment indicated an $NH_3 - N$ removal from 37.2 mg/l down to 34.8 mg/l. In all cases, the carbon column effluent had no chlorine residual.

Summary

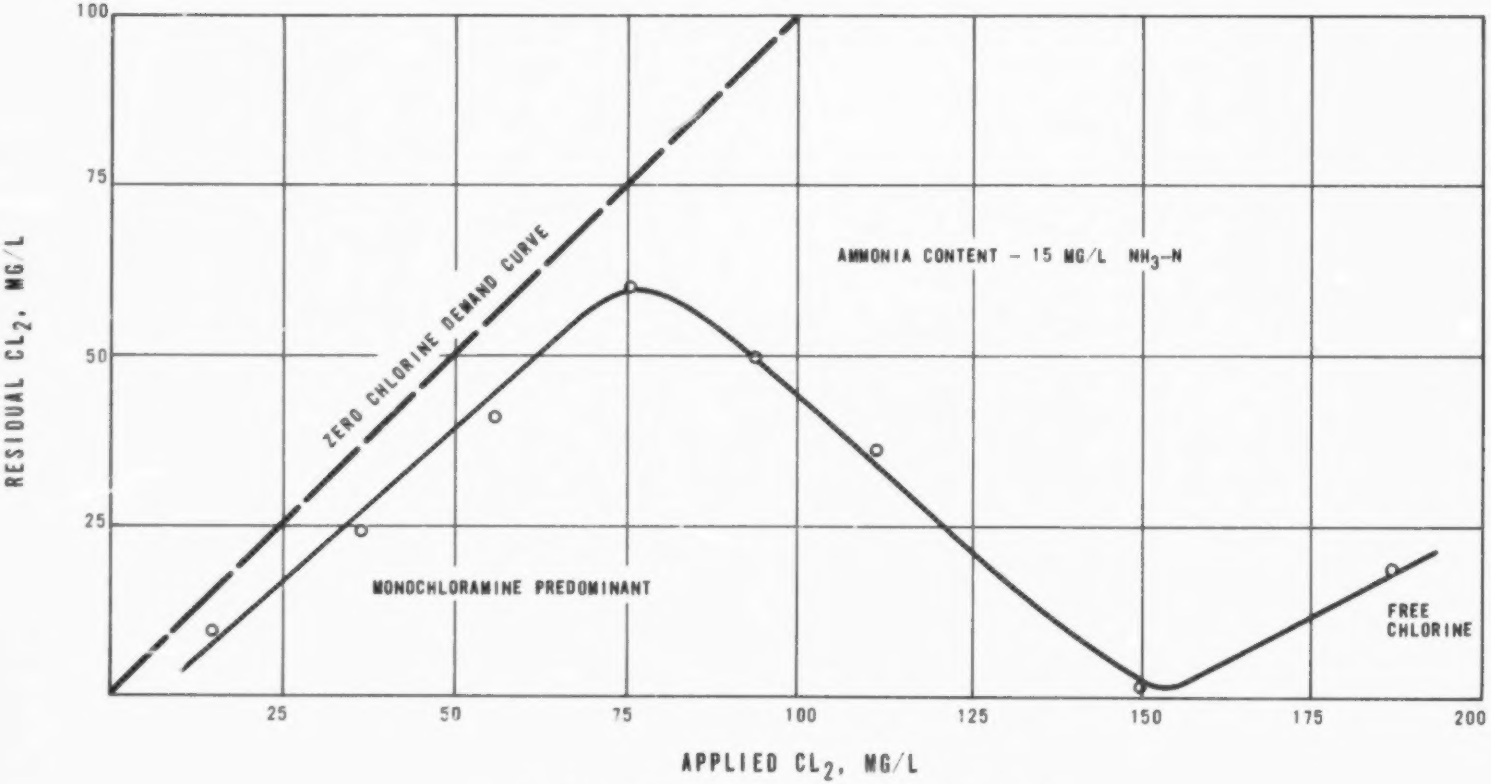
It appears the adsorption of both monochloramine and dichloramine to activated carbon is very minimal. Although the chlorine itself is removed, the ammonium ion passes through.

The only apparent method of ammonia removal by chlorination is by the oxidation of the chloramines to nitrous oxide, and the subsequent release as a gas. This is illustrated by the following equation:

$$NH_2Cl + NHCl_2 + HOCl \longrightarrow N_2O\uparrow + 4\,HCl$$

This would require chlorination to the breakpoint. Several investigators have indicated breakpoint chlorination requires approximately 10 mg of chlorine per mg of $NH - N$.(49,50) The breakpoint curve obtained during this study (Figure 93) substantiates this ratio.

Based on a chlorine cost of $0.07 per pound, the chemical cost of ammonia removal by breakpoint chlorination would be near $6 per million gallons per mg/l of $NH_3 - N$.



(CHLORINE SOURCE: SODIUM HYPOCHLORITE)

FIGURE 93

BREAKPOINT CHLORINATION CURVE FOR LIME-CLARIFIED SECONDARY EFFLUENT RECARBONATED TO A pH OF 9.0

CHAPTER VIII

LIME TREATMENT PROCESS PILOT PLANT

INTRODUCTION

Operating experience with the existing tertiary plant, along with the data collected on methods of nutrient removal, resulted in the development of a process which will remove phosphates, nitrogen, and organic matter, as well as enable recovery and reuse of chemicals used in the process. A schematic diagram of the pilot plant developed to investigate this process appears in Figure 94.

A 25-gpm capacity, five-chamber lime clarification unit (Figure 95) was designed and fabricated. Thus, a means was available of accurately evaluating lime dosages and various coagulant aids as to their effect on effluent quality of the lime clarification process. Also, the clarification unit supplied a continuous source of lime-clarified water to the pilot ammonia stripping tower. Pilot separation beds were employed to treat the tower effluent.

LIME CLARIFICATION PROCESS

Activated sludge secondary effluent was pumped through a flow totalizing meter to a flow rate control tank. Lime slurry was metered into the influent line, followed by polymer coagulant aids when used. The flow then passed to a rapid mix tank with a detention time of 2½ minutes and then on to a flocculation chamber. Slow agitation in the flocculation tank for approximately 5½ minutes, at 25 gpm, created floc particles of settleable size and maximized turbidity removal. The flocculated water was then passed to a circular settling basin with a detention period of near 1 hour. The floc settled to a conical sludge hopper for withdrawal. The clarified water flowed over a weir and continued to the final tank containing a sump pump, which pumped the water to the top of the pilot ammonia stripping tower.

AMMONIA STRIPPING TOWER

The pilot ammonia stripping tower to which the lime-treated water was pumped was described and pictured earlier in this report. The original 1½-inch by 2-inch packing consisted of untreated redwood slats supported at 30-inch centers. Through an extended period of use, the slats had weathered and sagged badly, causing poor flow distribution through the tower. This poor distribution eliminated adequate droplet formation, and thus further collection of meaningful ammonia removal data was impossible. The passing of lime-clarified water through the tower was continued, however, to investigate the possibility of a scale forming on the packing and inhibiting ammonia removal.

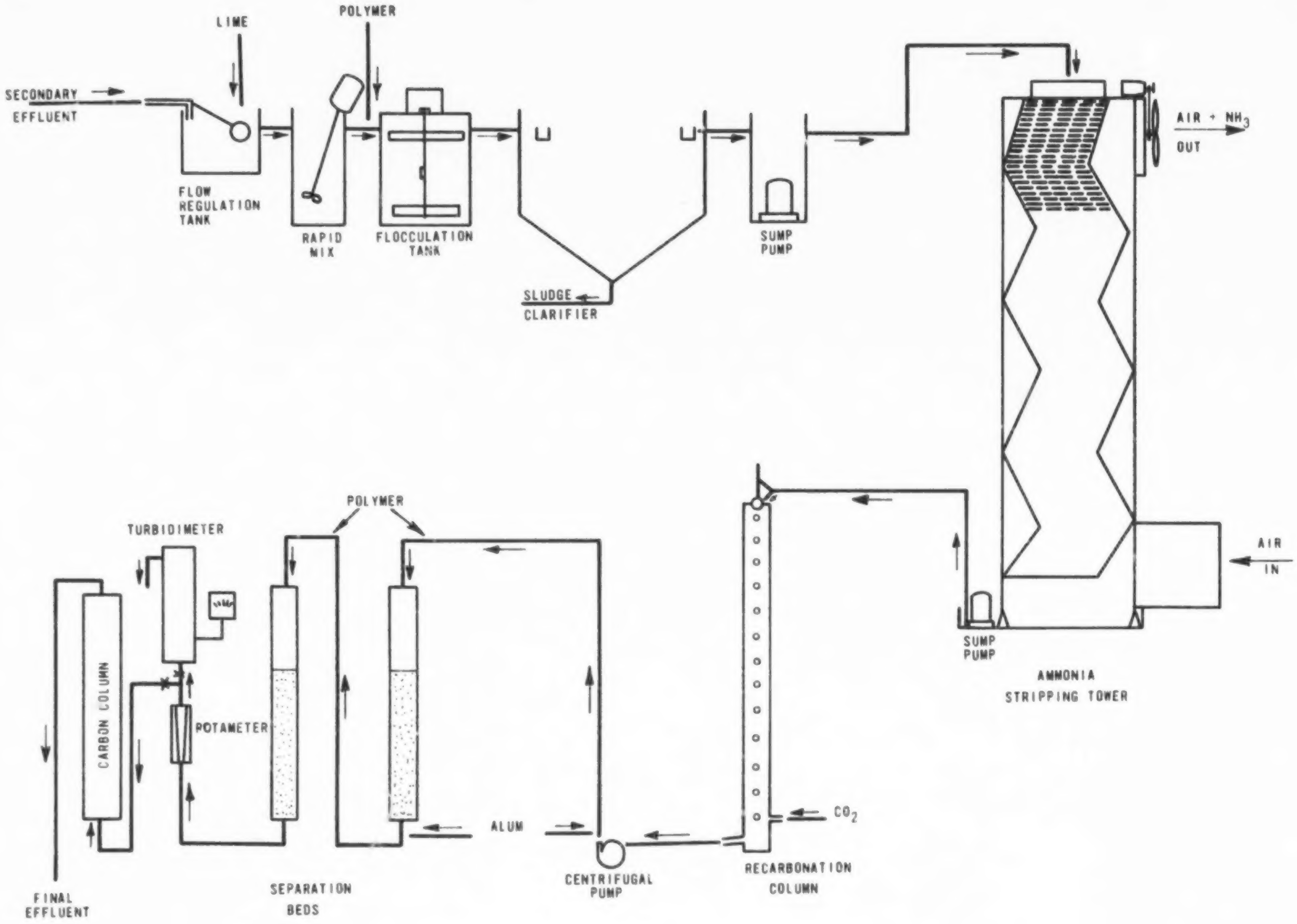
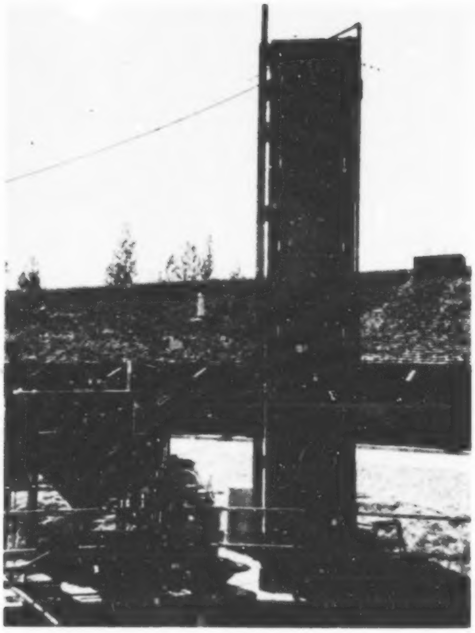
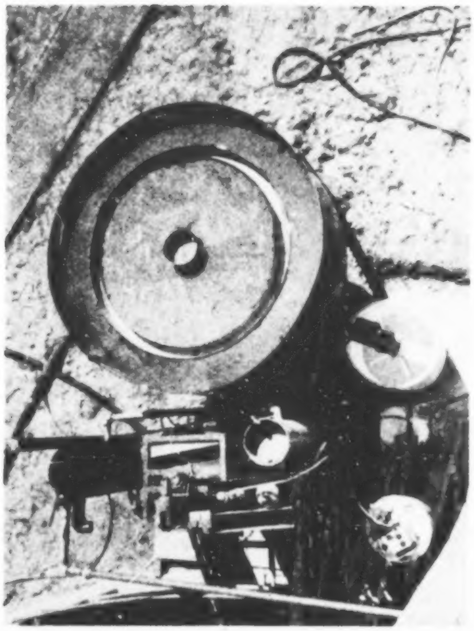

LIME
POLYMER
SECONDARY EFFLUENT
FLOW REGULATION TANK
RAPID MIX
FLOCCULATION TANK
SLUDGE
CLARIFIER
SUMP PUMP
AIR + NH_3 OUT
AMMONIA STRIPPING TOWER
AIR IN
SUMP PUMP
POLYMER
TURBIDIMETER
CARBON COLUMN
POTAMETER
CO_2
ALUM
FINAL EFFLUENT
SEPARATION BEDS
CENTRIFUGAL PUMP
RECARBONATION COLUMN

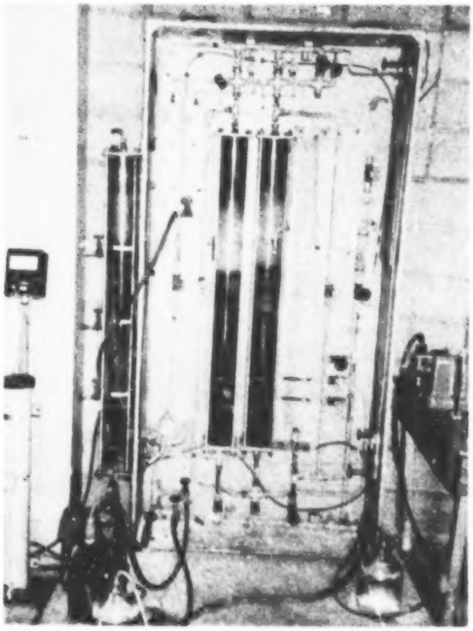
FIGURE 94
SCHEMATIC DIAGRAM OF
LIME TREATMENT PROCESS PILOT PLANT



LIME CLARIFICATION UNIT AND
AMMONIA STRIPPING TOWER



LIME CLARIFICATION UNIT



PILOT SEPARATION BEDS

FIGURE 95
LIME TREATMENT PROCESS PILOT PLANT

Over 160,000 gallons were passed through the tower during this period, and a crystaline deposit, rather than a tight scale, did form on the packing. However, this deposit was not judged to present a problem as it could be easily wiped off and did not build to the point that the surface texture of the wood was covered.

FILTRATION

A portion of the tower effluent was pumped to a recarbonation chamber. This device consisted of a 9-foot-deep section of 4-inch-diameter PVC pipe, through which CO_2 was bubbled countercurrent to the flow. Therein, the pH of the water was reduced to near 9.5 as the CO_2 gas affected the bicarbonate buffer system according to the following reactions:

$$CO_2 + 2OH^- \longrightarrow CO_3^= + H_2O$$

$$CO_3^= + Ca^{++} \longrightarrow CaCO_3 \downarrow$$

Further pH drop can be achieved by the addition of still more CO_2.

$$CO_2 + H_2O \rightleftharpoons H_2CO_3 \rightleftharpoons HCO_3^- + H^+$$

$$CO_2 + CO_3^= + H_2O \longrightarrow 2\,HCO_3^-$$

The recarbonated water was then pumped to small Microfloc separation beds, pictured in Figure 95, with media similar to that in the main plant. At this point, residual turbidity from the lime clarifier as well as calcium carbonate floc created during the recarbonation process, were removed by coagulation and filtration in the multimedia beds.

CARBON ADSORPTION

The addition of an upflow carbon column in series with the separation beds made it possible for the entire treatment process of the South Lake Tahoe Water Reclamation Plant to be studied on a pilot scale. However, it was elected not to study any effects on the carbon treatment portion during this period.

DATA COLLECTED

Data collected to date from operation of the pilot plant are presented in Table 27.

During the test period, the secondary effluent frequently contained a high concentration of carryover activated sludge, as is indicated by the variability in the turbidity. The addition of lime, in most cases, created a settleable floc, although the effluent turbidity varied considerably. Phosphate reduction averaged approximately 90 percent through the clarifier. The addition of certain anionic polymers was occasionally helpful in maximizing floc size, but this did not aid in phosphate removal or turbidity reduction. The volume of the resultant lime sludge averaged approximately 2 percent of the throughput volume, and the average sludge density was approximately 2.5 percent solids.

The pH resulting from the lime addition was more than adequate for ammonia stripping. The stripping tower packing was replaced in April 1967, with the packing and support system tentatively chosen for the plant-scale tower. This consisted of a 1-1/2-inch by 2-inch packing of fir lath. The lath was pressure-treated with copper arsenite and supported at 22-inch centers. The data collected from operation after the tower was repacked, mostly concerned with evaluating the temperature effect, were presented in the chapter on nitrogen removal.

Table 27

LIME TREATMENT PILOT PLANT OPERATIONAL DATA

General Operating Conditions

Lime Dosage – 500 mg/l $Ca(OH)_2$
Stripping Tower Loadings:
  Liquid – 0.7 to 2.0 gpm/sq ft
  Air – 500 cu ft/gal
Air Temperature – 23 to 45°F
Alum Dosage – 20 to 50 mg/l
Coagulant Aid – 0.2 to 0.5 mg/l anionic
Filtration Rate – 5.0 gpm/sq ft

Operating Efficiency

	Sampling Point				
Parameter	Secondary Effluent	Clarifier Effluent	Stripping Tower Effluent	Recarbonated Effluent	Separation Bed Effluent
pH	7.0–7.4	11.4–12.0	11.0–11.8	9.4–9.6	9.2–9.4
Temp., °C	11–17	10–17	0–12	14–15(1)	14–15(1)
Turbidity	35–>100	5–35	–	>100 ($CaCO_3$ floc)	1.0–3.5
NH_3-N, mg/l	15–35	15–35	5–12(2)	–	5–12(2)
PO_4, mg/l	15–35	0.8–3.0	–	–	0.8–2.0
Ca, mg/l	8–14	56–104	–	–	35–55

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(1) This portion of pilot plant located indoors – tower effluent reheated.

(2) Temperature dependent.

Jar tests were run on recarbonated tower effluent to evaluate various coagulant aids. The results are presented in the appendix of this report. These results show the slightly anionic to anionic polymers to be the most effective aid in maximizing floc formation. Other tests indicate the provision of a 10-minute slow mix period following recarbonation will, at times, create a settleable floc without use of a polymer. The cationic aids have yet to be evaluated for use as filter conditioners.

Calcium recovery was highly dependent upon the recarbonation pH level. During this study period, a pH of 9.5 was the attempted recarbonation level; but even at this level, calcium losses were higher than previously achieved, as 35 to 55 mg/l of calcium in solution were lost through the filters. Effluent phosphate concentrations were also higher than anticipated, as were effluent turbidities.

The above results prompted the development of a plan involving two-stage recarbonation. During the first stage, the tower effluent would be recarbonated to a pH of only approximately 9.3, thereby minimizing calcium losses to solution. This would be followed by a reaction period in which the calcium carbonate floc would be formed A second recarbonation to about pH 8.5 would prepare the water for further treatment and discharge.

CHAPTER IX

ALGAE GROWTH STUDIES WITH RECLAIMED WATER

INTRODUCTION

A considerable amount of experimental work was put into the development of a tertiary treatment for the removal of organic matter and critical nutrients. In order to obtain a visual representation of the results such a treatment process might have on inhibiting the algae growth potential of waste water, a limited algae study was instigated.

EXPERIMENTAL PROCEDURES AND RESULTS

A series of gallon glass jugs were used to store (a) samples of waste water after various stages of treatment, and (b) samples of Lake Tahoe water to which various concentrations of phosphorus and nitrogen compounds had been added. Phosphorus was added in the form of KH_2PO_4, while nitrogen was added as KNO_3. The concentrations of phosphorus and nitrogen chosen were based both upon reported limiting concentrations of each,(18) as well as technical ability to achieve these concentrations in reclaimed water.

Table 28 presents the contents of the samples, as well as a visual observation of the growth of algae after approximately 2½ months of storage. The samples were stored indoors and were subject to both artificial and natural light.

As is evident from Table 27 , algae growth in these samples diminished with higher degrees of treatment. A trace of algae growth was appearing in the final tertiary effluent, which had received alum coagulation, filtration, and carbon adsorption, and which contained approximately 0.1 mg/l PO_4. A mixture of algal species was present, the only common one readily identifiable being anabaena. At the time of observation, no algae was noticeable in any of the lake water samples.

Table 28

OBSERVATIONS ON ALGAE GROWTH STUDIES

Sample No.	Contents	Treatment	Algae Growth Observations
1	Primary Effluent	Plain Sedimentation	Large Amount
2	Secondary Effluent	Activated Sludge	Considerable
3	Chemical Clarifier Effluent	Alum Coagulation	Very Noticeable
4	Separation Bed Effluent	Alum Coagulation and Filtration	Noticeable
5	Tertiary Effluent	Alum Coagulation, Filtration, Carbon Adsorption	Trace
6	Lake Tahoe Water(Blank)	-	Not Noticeable
7	Lake Tahoe Water +.05 mg/1 P	-	Not Noticeable
8	Lake Tahoe Water + 0.3 mg/1 N	-	Not Noticeable
9	Lake Tahoe Water + .05 mg/1 P+0.3 mg/1 N	-	Not Noticeable
10	Lake Tahoe Water + .01 mg/1 P	-	Not Noticeable
11	Lake Tahoe Water + 1:1,000 of Tertiary Effluent	-	Not Noticeable

CHAPTER X

DESCRIPTION AND DESIGN OF THE EXPANDED 7.5-MGD SOUTH LAKE TAHOE RECLAMATION PLANT

Demonstration grant laboratory and pilot plant studies resulted in the development of a design for the current expansion of the South Tahoe P. U. D. Water Reclamation Plant to a capacity of 7.5 mgd. Figure 96 is a schematic diagram for the expanded plant. A more complete description appeared in a report by Culp. (51)

PROCESS DESCRIPTION

Of course, the existing primary and secondary facilities had to be expanded to meet the additional capacity requirements. Incorporated in the expansion is a sludge-thickening and incineration process for handling the primary and waste activated sludge. The first major addition to the existing plant with respect to tertiary treatment was the incorporation of a chemical flocculation chamber and clarifier following the secondary clarifiers. Here, the suspended matter in the secondary effluent is coagulated and removed, and the majority of the phosphates are precipitated.

The high pH (11.5 - 12.0) effluent from the chemical clarifier will be pumped to a lath-packed stripping tower in which the ammonia-nitrogen will be removed. The substructure of this tower will contain facilities for recirculation, effluent recarbonation, and calcium carbonate recovery. First-stage recarbonation to a pH of about 9.3 will occur as the waste water leaves the recirculation launder. This will be accomplished by means of CO_2 obtained from sludge incineration and recalcination furnace stack gases. Following the formation of $CaCO_3$, the water will be further recarbonated to a pH near 8.5.

Following second-stage recarbonation, the water will pass to the expanded version of the original tertiary treatment plant. The separation beds will provide coagulation-filtration of $CaCO_3$, residual turbidity, and phosphates. The final treatment will still consist of additional organic removal and polishing by granular activated carbon.

LIME FEEDING AND RECOVERY

As mentioned earlier, the key to lime recovery without buildup of phosphates in the system is the split-feed lime application.

In general, the design of lime-feeding equipment at South Tahoe is as follows: The initial lime dosage is applied to the primary clarifiers; the lime used for this feed will be entirely recalcined lime, which carries with it a

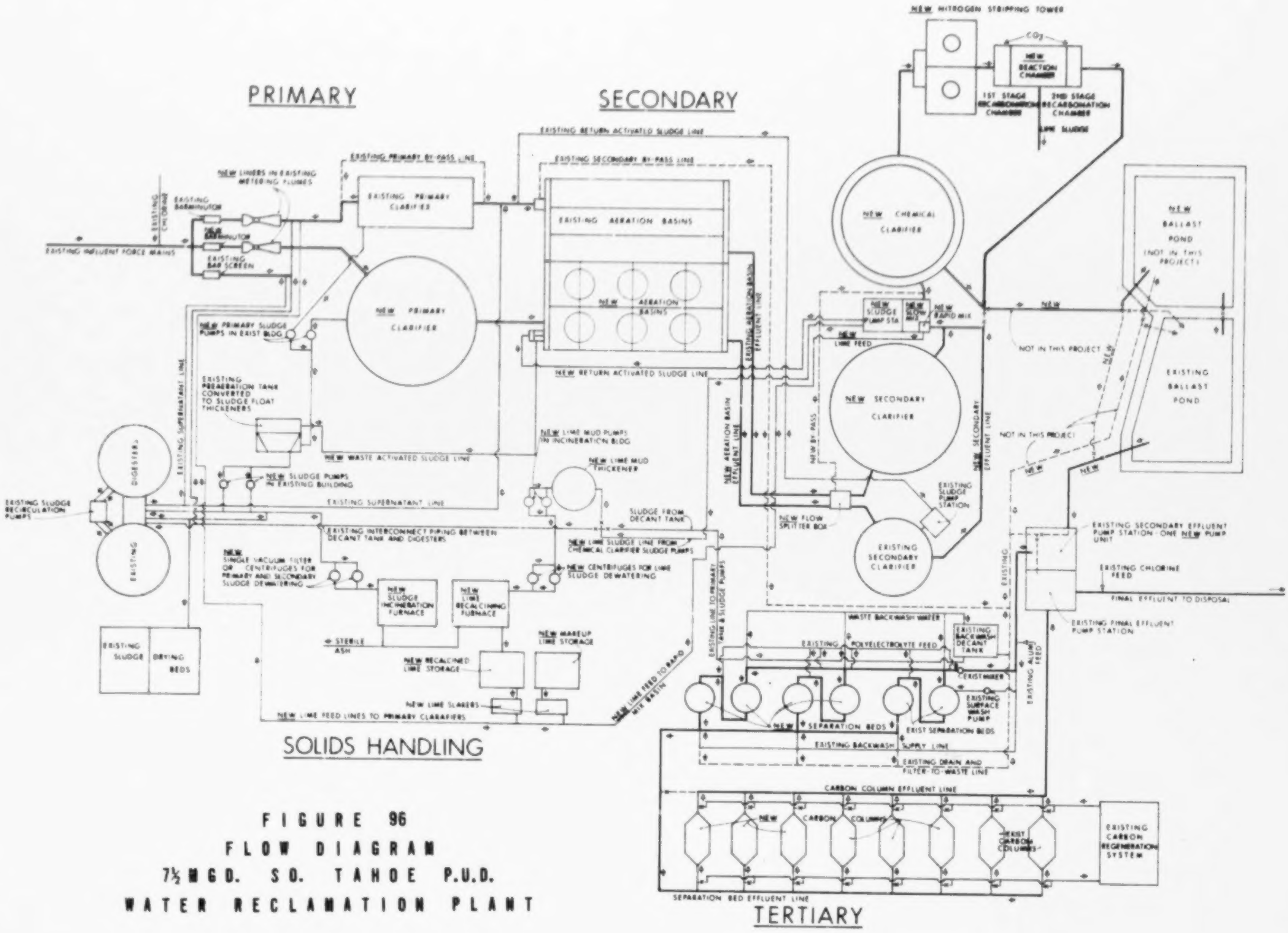
PRIMARY
SECONDARY
SOLIDS HANDLING
TERTIARY
EXISTING INFLUENT FORCE MAINS
EXISTING CHLORINE
EXISTING BARMINUTOR
NEW BARMINUTOR
EXISTING BAR SCREEN
NEW LINERS IN EXISTING METERING FLUMES
EXISTING PRIMARY BY-PASS LINE
EXISTING PRIMARY CLARIFIER
NEW PRIMARY CLARIFIER
NEW PRIMARY SLUDGE PUMPS IN EXIST BLDG
EXISTING PREAERATION TANK CONVERTED TO SLUDGE FLOAT THICKENERS
EXISTING SUPERNATANT LINE
NEW WASTE ACTIVATED SLUDGE LINE
NEW SLUDGE PUMPS IN EXISTING BUILDING
EXISTING SLUDGE RECIRCULATION PUMPS
DIGESTERS
EXISTING
EXISTING SLUDGE DRYING BEDS
EXISTING RETURN ACTIVATED SLUDGE LINE
EXISTING SECONDARY BY-PASS LINE
EXISTING AERATION BASINS
NEW AERATION BASINS
EXISTING AERATION BASIN EFFLUENT LINE
NEW RETURN ACTIVATED SLUDGE LINE
NEW AERATION BASIN EFFLUENT LINE
NEW BY-PASS
NEW LIME MUD PUMPS IN INCINERATION BLDG
NEW LIME MUD THICKENER
SLUDGE FROM DECANT TANK
NEW LIME SLUDGE LINE FROM CHEMICAL CLARIFIER SLUDGE PUMPS
EXISTING INTERCONNECT PIPING BETWEEN DECANT TANK AND DIGESTERS
NEW SINGLE VACUUM FILTER OR CENTRIFUGES FOR PRIMARY AND SECONDARY SLUDGE DEWATERING
NEW CENTRIFUGES FOR LIME SLUDGE DEWATERING
NEW SLUDGE INCINERATION FURNACE
NEW LIME RECALCINING FURNACE
STERILE ASH
NEW MAKEUP LIME STORAGE
NEW RECALCINED LIME STORAGE
NEW LIME SLAKERS
NEW LIME FEED LINES TO PRIMARY CLARAFIERS
NEW LIME FEED TO RAPID MIX BASIN
EXISTING LINE TO PRIMARY TANK & SLUDGE PUMPS
NEW NITROGEN STRIPPING TOWER
CO2
NEW REACTION CHAMBER
1ST STAGE RECARBONATION CHAMBER
2ND STAGE RECARBONATION CHAMBER
LIME SLUDGE
NEW CHEMICAL CLARIFIER
NEW SLUDGE PUMP STA
NEW SLOW MIX
NEW RAPID MIX
NEW LIME FEED
NEW SECONDARY CLARIFIER
NEW SECONDARY EFFLUENT LINE
NEW FLOW SPLITTER BOX
EXISTING SLUDGE PUMP STATION
EXISTING SECONDARY CLARIFIER
NEW
NEW BALLAST POND (NOT IN THIS PROJECT)
EXISTING BALLAST POND
NOT IN THIS PROJECT
NEW
NOT IN THIS PROJECT
NEW
NEW
EXISTING
EXISTING SECONDARY EFFLUENT PUMP STATION - ONE NEW PUMP UNIT
EXISTING CHLORINE FEED
FINAL EFFLUENT TO DISPOSAL
EXISTING FINAL EFFLUENT PUMP STATION
WASTE BACKWASH WATER
EXISTING POLYELECTROLYTE FEED
EXISTING BACKWASH DECANT TANK
EXIST MIXER
EXISTING ALUM FEED
EXISTING SURFACE WASH PUMP
NEW SEPARATION BEDS
EXIST SEPARATION BEDS
EXISTING BACKWASH SUPPLY LINE
EXISTING DRAIN AND FILTER-TO-WASTE LINE
CARBON COLUMN EFFLUENT LINE
NEW CARBON COLUMNS
EXIST CARBON COLUMNS
EXISTING CARBON REGENERATION SYSTEM
SEPARATION BED EFFLUENT LINE

FIGURE 96
FLOW DIAGRAM
7½ MGD. SO. TAHOE P.U.D.
WATER RECLAMATION PLANT

percentage of tricalcium phosphate. This $Ca_3(PO_4)_2$, along with any additional $Ca_3(PO_4)_2$, precipitated by the initial lime dosage, is collected with primary sludge and is eventually disposed of as sludge incinerator ash. Thus, this $Ca_3(PO_4)_2$ leaves the system at this point and does not report back as a recycling load to the lime reclamation equipment. The initial portion of the split feed is controlled by the pH of aeration basin influent to a pH not detrimental to the activated sludge process. The remainder of the required lime dosage is made up of the balance of available reclaimed lime, plus the required amount of virgin or makeup lime, and is applied to the sewage at the flash-mix flocculation basin. pH control of dosage is again employed.

Calcium sludge is to be pumped from the chemical clarifier, from the recarbonation sedimentation basin, and from the separation bed decant tank to a spent lime mud thickener. The thickened lime sludge will then be centrifuged and passed to a multiple hearth furnace for recalcination.

OPERATION TO DATE

To date, the only portion of the expanded treatment plant which has been completed is the chemical clarifier. This has been put into operation using alum as the coagulant, thereby reducing the load on the separation beds. A high quality chemical clarifier effluent has been obtained at present loadings. Typical analysis includes a turbidity of 0.5 to 3.0 (Formazin standard units), a phosphate concentration of 0.5 to 1.5 mg/l PO_4, and a COD of 20 to 40 mg/l.

ESTIMATED COST OF TERTIARY PROCESS

The estimated costs, totaling approximately $150/mg, shown in Table 29 are those anticipated to provide phosphate, nitrogen, and organic removal, as well as recovery and reuse of coagulant for a 7.5-mgd plant at the South Tahoe Public Utility District.

Capital costs are amortized at 4 percent interest over 20 years. Chemical costs are f.o.b., South Lake Tahoe and include 4 percent California sales tax. Power is based upon a demand charge of $2.30 per kw hour per month and an energy charge of $0.008 per kw hour. Maintenance is based upon an annual cost of 3 percent of the equipment investment and labor costs of $3.75 per hour.

The capital cost of the chemical clarifier includes the expenditures for not only the clarifier, but the recarbonation system, lime handling and storage facilities, slakers, feeders, and flash mix, and flocculation basin. The lime cost is based upon 25 percent makeup, as explained earlier.

Table 29

ESTIMATED COST OF PHOSPHATE, NITROGEN, AND ORGANIC REMOVAL AT 7.5-MGD SOUTH TAHOE PUBLIC UTILITY DISTRICT WATER RELCAMATION PLANT

Chemical Clarifier & Recarbonation System		Cost Dollars/Million Gallons	
Capital		$11.75	
Lime - 400 mg/l as CaO			
Reclaimed	31.80		
Makeup	10.90		
		42.70	
Power		5.35	
Maintenance		1.90	
Labor		2.00	
			$63.70
Ammonia Stripping - 95% Removal of Ammonia (Summer Operating Conditions)			
Capital		4.60	
Power		7.60	
Maintenance		0.25	
Labor		2.00	
			14.45
Separation Beds			
Capital		17.30	
Chemicals		9.90	
Power		7.55	
Maintenance		2.20	
Labor		4.00	
			40.95
Activated Carbon			
Capital		18.00	
Regeneration		11.75	
Maintenance		1.10	
			30.85
Total Estimated Capital and Operating Costs Phosphate, Nitrogen, and Organic Removal, Plus Recovery and Reuse of Lime			$149.95

CONCLUSION

The culmination of the past 2 years of demonstration grant research will be a 7.5-mgd water reclamation plant capable of removing phosphates, nitrogen, suspended matter, and organic pollutants present in the waste water. Moreover, the economical recovery and reuse of a large portion of the chemicals used will reduce operating costs as well as diminish the problem of the disposal of large quantities of chemical sludge. Table 30 indicates the anticipated quality of the final effluent from this plant.

Table 30

ANTICIPATED EFFLUENT QUALITY FROM THE
SOUTH TAHOE WATER RECLAMATION PLANT

Parameter	Raw Sewage		Final Effluent
BOD mg/l	200-400		<1
COD mg/l	400-600		3-25
Total Organic Carbon mg/l	high		<1-7.5
ABS mg/l	2-4		<0.01-0.5
Color, units	high		<5
Turbidity SJU	high		<0.5
PO$_4$ mg/l as PO$_4$	25-30		0.2-1
Nitrogen			
Organic N mg/l as N	10-15		0.3-2.0
Ammonia N mg/l as N	25-35	Summer	0.3-1.5
		Winter	<10
Nitrate and Nitrite mg/l as N	0		0
pH	7.2-7.4		8.3-8.6
Alkalinity mg/l as CaCO$_3$	110-130		250
Sulphate mg/l	20		20

The plant effluent will be pumped 26 miles through an export line into Alpine County, California. There, a multi-purpose storage reservoir will be constructed which will be filled primarily with effluent from the water reclamation plant. Current uses planned for the reservoir include both recreation and irrigation of grassland.

CHAPTER XI

LIST OF PUBLICATIONS
RESULTING FROM DEMONSTRATION GRANT PROGRAM

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APPENDIX A

EVALUATION OF POLYMERS

Table A-1

WATER — EFFLUENT FROM LIME CLARIFIER, LIME DOSAGE — 500 mg/l $Ca(OH)_2$

Recarbonation — pH 9.0
Turbidity — Initial, 65 — $CaCO_3$ Floc Very Visible [pH 7.0 Turb. < 2.5]
Alum — 50 mg/l on All Tests
Rapid Mix — 30 Sec. (100 rpm)
Flocculation — 8 Minutes (35 rpm) — Settle — 5 Minutes

I. CALGON ST. 270 (SLIGHTLY ANIONIC)

Dose	Floc. Obs.	Turb.	pH
0	fair good	1.8	8.6
0.2	very good	1.9	8.6
0.4	exc.	1.8	
0.6	exc.		8.8
0.8	exc.		
1.0	exc.	2.1	8.8

II. HERCULES A-5 (ANIONIC)

Dose	Floc. Obs.	Turb.	pH
0	fairly good	1.4	8.7
0.2	exc.	1.6	8.7
0.4	very good		
0.6	exc.	1.8	8.7
0.8	exc.		
1.0	exc.	1.8	8.7

III. CALGON 230 (NONIONIC)

Dose	Floc. Obs.	Turb.	pH
0	fairly good	2.1	8.6
0.2	fairly good	1.8	8.7
0.4	fairly good		
0.6	good	1.6	8.7
0.8	good		
1.0	(little difference)	1.4	8.7

(Cont'd)

Table A-1 (Cont'd)

IV. MAGNIFLOC 990 (NONIONIC)

Dose	Floc. Obs.	Turb.	pH
0	fairly good	1.6	8.7
0.2	fairly good	1.6	8.7
0.4	fairly good		8.8
0.6	fairly good	1.6	8.8
0.8	no difference		
1.0	no difference	1.8	8.9

V. DREWFLOC 31 (NONIONIC)

Dose	Floc. Obs.	Turb.	pH
0	fairly good	1.7	8.6
0.2	fairly good	1.6	8.6
0.4	fairly good		
0.6	fairly good	1.7	8.7
0.8	little change		
1.0	little change	1.7	8.7

VI. HERCULES A-1 (ANIONIC)

Dose	Floc. Obs.	Turb.	pH
0 mg/l	fairly good	1.8	8.7
0.2	exc.	2.0	8.8
0.4	exc.		
0.6	exc.	2.3	8.8
0.8	exc.		
1.0	exc.	2.3	8.8

VII. NALCO 675 (ANIONIC)

Dose	Floc. Obs.	Turb.	pH
0	fairly good	1.8	8.6
0.2	good	2.3	8.7
0.4	good		
0.6	good	2.4	8.8
0.8	very good	2.6	8.8
1.0	exc.	2.5	8.8

(Cont'd)

Table A-1 (Cont'd)

VIII. DREWFLOC 31 – NEW MIXTURE – 0.1% BY VOLUME STOCK SOLUTION

Dose	Floc. Obs.	Turb.	pH
0 mg/l	fair	2.8	8.5
0.2 (mls)	fairly good	2.2	8.7
0.4	fairly good		
0.6	fairly good	2.2	8.8
0.8	fairly good		
1.0	good	2.0	8.8

IX. NALCO 603 (CATIONIC)

Dose	Floc. Obs.	Turb.	pH
0	fairly good	2.1	8.7
0.2	fairly good	2.4	8.7
0.4	fairly good		
0.6	fairly good	2.7	8.8
0.8	fairly good		
1.0	no change	2.9	8.8

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